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A RAPID CAPITAL COST ESTIMATION METHOD FOR NATURAL GAS PROCESSING PLANTS

by

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A THESIS

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The undersigned certify that they have read, and recommend to the Faculty of Graduate Studies for acceptance, a thesis entitled A RAPID CAPITAL COST ESTIMATION METHOD FOR NATURAL GAS PROCESSING PLANTS submitted by David Grant Gunderson in partial fulfilment of the requirements for the degree of Master of Business Administration.

ABSTRACT

Alberta's natural gas processing industry plays an increasingly important role in our economy, and the design and construction of the required processing facilities has become a major segment of the petroleum industry.

Cost estimation procedures in this field are inadequate except at the detailed design level. There is a definite need for a rapid, accurate method of cost estimation for use at the pre-design stage.

This study assumes that most natural gas processing plants can be broken down into one or more processing sections: (1) the natural gas liquids recovery section, (2) the acid gas removal section, and (3) the sulphur recovery section.

Information on capital costs and operating parameters for the sixteen plants studied was obtained from the files of Pacific Petroleums Ltd., the company that employs the writer. The capital cost of each existing section of these plants was plotted against a function containing the significant operating parameters affecting the cost of the section. The result was a logical set of capital cost curves which allow rapid estimation of capital cost with reasonable accuracy.

ACKNOWLEDGEMENTS

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CHAPTER I

INTRODUCTION

Natural gas has become one of Alberta's most important assets. The major oil companies in this province are spending considerable time and effort in searching for and marketing this product.

Most Alberta natural gas requires some form of processing before it can be marketed. This processing involves the removal of certain components from the gas, because of gas quality or economic considerations.

The processing facility is usually a significant part of any gas marketing project, and its cost can have profound effects on the economic viability of the project. This cost can range from almost nothing where the gas occurs naturally at nearly marketable specifications, to tens of millions of dollars where massive quantities of by-products and contaminants must be removed.

With the increasing importance of natural gas processing plants in today's petroleum industry, it would be expected that some reasonably accurate methods of estimating the capital costs of these plants would be present. Unfortunately this is not so. Presently the only type of estimate consistently providing a satisfactory degree of accuracy is the detailed cost estimate provided by the processing plant construction contractors <u>after</u>

they have completely designed the plant. It should be evident to the reader that before the gas marketing project reaches the stage of soliciting bids from contractors, at least an approximate estimate of the required capital expenditures is necessary for the company contemplating the project.

I. THE PURPOSE OF THIS STUDY

The purpose of this study was to provide the management and engineering staff of an oil company with a satisfactory short method of estimating the capital costs of natural gas processing plants. The required method was to be rapid and easily understood. Initial discussions with process engineers indicated that a satisfactory accuracy was approximately plus or minus twenty per cent.

It was not intended that the method presented here replace the detailed cost estimate provided by the contractors. In this regard it could only be used as a rough check on the bids received. However, much economic planning and decision making is normally accomplished prior to the solicitation of bids, and the method described in this paper was designed to assist the decision maker at the earlier stages of the project.

II. THE PRESENT STATE OF KNOWLEDGE ON GAS PLANT CAPITAL COST ESTIMATING

Very little literature was available on gas plant capital

cost estimation procedures. The most complete coverage of the topic was done recently by the Petroleum Industry Training Service in their 1967 edition of "Gas Technology".[1] This paper provided some very general rules for cost estimation procedures and discussed the important factors to be considered. There was little mention of the possibility of developing a satisfactory overall estimation method for use in the pre-design stage, and discussions with the engineers involved in the publication suggested that because of the complexities involved no such method was possible.

This attitude was confirmed in discussions with process engineering consultants in Calgary. Some consultants had attempted the derivation of such a method but had little success.

Most of the other literature examined was presented on the same level as "Gas Technology". A great deal of discussion was offered on the general factors to consider in cost estimating, and the accent was placed on the detailed cost estimate. The problem of the preliminary estimate was usually ignored except for references to one specific "rule of thumb" common to the industry.[2]

There were a few articles presented in the various oil industry publications that did attempt to solve the problem of the preliminary cost estimate. Generally, these attempts suffered from two faults. First, they did not go far enough as the methods

proposed were applicable to only a small segment of the total processing plant. Second, there was almost no statistical support for the majority of the articles, which often started with a plant of known cost and used the "rule of thumb" referred to above as a major parameter.

The one notable exception to these comments dealt in packaged sulphur recovery plants. Black, Sivalls and Bryson Limited, of Edmonton, provided an excellent chart of plant capital cost versus sulphur production. Although the chart was used for advertising purposes, the data that it contained was the result of actual bids tendered by the company, as well as price research on competitors' projects. This writer discussed the data thoroughly with Black, Sivalls and Bryson personnel, and a suitably modified version of their chart was used in this paper.

The prevailing attitude of the industry parallels the lack of progress found in the literature. The stress has been placed on the detailed cost estimate because no suitable substitute has been found. Although computer techniques are slowly solving the problems of rapid detailed design and cost estimation, it will be many years before the need for a good preliminary design method is alleviated. Those who are required to provide preliminary estimates often request the services of a consultant, which can be expensive and time consuming. Experienced gas process engineers cannot usually

be relied upon to make a reasonable estimate of the capital expenditures required for a specific processing plant. The writer has observed, on four different occasions, the estimating procedures followed at the managerial level. The error (compared with the actual plant cost or at least the final detailed estimate) ranged from minus thirty per cent to plus three hundred per cent. It was after observing these difficulties that the writer began to investigate the problem.

III. SOURCES OF DATA

The prime source of data for this study was provided by the files of Pacific Petroleums Ltd., the company that employs the writer. This of course limited the study to those plants in which the company had an interest. Efforts to gain access to information on other processing plants met with little success because the managements of other companies saw no benefits in releasing information to a competitor.

"Canadian Petroleum", an industry periodical, publishes certain specifications on all Canadian gas processing plants.[3] Unfortunately, this data was found to be very unreliable, particularly in the capital costs quoted. This precluded the use of the information provided, except for comparison purposes.

The writer would be remiss if he did not mention the

substantial information collected during the course of many discussions with company personnel. Most of the data collected was subjected to these discussions and much essential information was provided in this way.

IV. OUTLINE OF THE THESIS

Chapter II outlines the functions and operations of the various segments of a natural gas processing plant. It is felt that this chapter, together with the definitions provided at the end of Chapter I, will enable the non-technical reader to understand any theoretical discussions that are presented to support the statistical relationships put forward in the later chapters.

Chapter III investigates the most common estimating procedures now in industry use. Particular attention is paid to the detailed cost estimate and therefore, of necessity, the actual design of the gas plant. The much used "point six" rule is discussed, along with the other rapid estimating procedures. Accuracy ranges for all these estimating procedures are provided.

Chapter IV is the main body of this study. It is a statistical analysis of the capital cost as a function of certain operating parameters of natural gas processing plants. The chapter first sets the limitations on the scope of the study and discusses the data collection procedure used. Of special significance is the derivation of the price

index used to convert the costs of the plants studied into dollars of uniform purchasing power.

Significant correlations are suggested for all of the major segments of the gas plants studied. Theoretical discussions are provided in support of these correlations.

Chapter V summarizes the capital cost estimation results for all plants included in the study. The estimates are compared with actual costs and the resulting errors are calculated and discussed. The chapter closes with a general discussion of the results and conclusions of the study.

V. DEFINITIONS USED IN THIS STUDY

The following definitions, although common to the industry, are provided to assist the non-technical reader in understanding this paper.[4]

Natural gas. This is a naturally occuring complex mixture of hydrocarbon and non-hydrocarbon constituents existing as a vapor at room temperature.

Well effluent. This is the untreated fluid (gas or liquid) from the reservoir.

Raw gas. This is the untreated or slightly treated gas which the plant is designed to process.

Pipeline gas or sales gas. This is natural gas with a quality suitable for domestic fuel or for meeting the pipeline companies' quality specifications.

Sour gas. This is natural gas that contains a significant quantity of hydrogen sulphide or carbon dioxide.

Acid gas. This is defined as that portion of the natural gas which is made up of hydrogen sulphide and carbon dioxide. Therefore, all sour gas contains acid gas.

Sweet gas. This is natural gas that is essentially free of acid gas.

Natural gas liquids (N.G.L.). These are liquid hydrocarbons recovered from the raw gas stream by the processing facility. They are made up of propane, butanes and condensate. Often the term is used to mean only the propanes and butanes, with the condensate being allotted a class by itself.

Condensate. This is hydrocarbon liquid consisting of pentanes and any hydrocarbons having a higher atomic weight than pentanes. It is sometimes referred to as "pentanes plus".

Pipeline specification gas. The specifications required by Trans Canada Pipelines Limited will be taken as representative of the industry requirements.

Gas plant capital cost. The capital costs investigated by this study include only those costs pertaining to the processing plant itself. The costs of gas gathering facilities, gas compression facilities, product loading facilities, and product transportation facilities are not included. Land costs, although usually insignificant, are also excluded.

Auxiliary items such as storage facilities, plant buildings, safety equipment, site roads, and utility installations are included in this study.

This study also excludes highly automated plants designed for unattended operation. This type of installation is not yet common in Alberta and is of course more expensive than a conventionally operated plant.

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[1]Gas Technology (Calgary, Petroleum Industry Training Service, 1968) This was the text provided for a course in gas processing attended by the writer.

[2]This is the "point six" rule commonly used in the industry. It is discussed fully in Chapter III.

[3]"CPE's 5th Census of Canadian Gas Processing Plants", Canadian Petroleum, July 1967, pp. 64-66.

[4]Petroleum Industry Training Service, op. cit. p. 1-3.

CHAPTER II

THE FUNCTION AND OPERATION OF NATURAL GAS PROCESSING PLANTS

I. FUNCTION

The prime function of most natural gas processing plants is to purify the raw natural gas as it arrives at the plant inlet in order that the gas can meet the rigid quality specifications laid down by the pipeline companies and utilities purchasing the gas. Although some gas wells can produce gas meeting purchaser specifications, the majority of natural gas produced in Alberta requires processing before marketing.

In some processing plants, the elements removed in the purification process become the most economically important product. A good example of this is the production of sulphur in today's market. Free World sulphur prices have more than quadrupled in the last six years, resulting in a proliferation of sulphur recovery plants in Canadian gas fields.

Natural gas, as produced by the well, can have the following components, in amounts varying from "trace" to large percentages:[1]

TABLE I

TYPICAL COMPONENTS OF NATURAL GAS

Hydrocarbons*	Non-Hydrocarbons**
Methane (C_1)	Water (H_2O)
Ethane (C_2)	Hydrogen Sulphide (H_2S)
Propane (C_3)	Carbon Dioxide (CO_2)
Butanes (C_4)	Nitrogen (N_2)
Pentanes and heavier hydrocarbons (C_{5+})	Oxygen (O_2)
	Helium (He)

*Accepted abbreviated chemical formula.
**Chemical formula.

The hydrocarbons listed in Table I are in order of increasing atomic weights, with the lightest component (methane) being the prime required constituent of the gas sold to the pipeline companies and utilities. Ethane, although present in much smaller quantities, also remains in the gas phase and is sold along with the methane. Next in order are: propane (C_3), butanes (C_4), and finally condensate or pentanes plus. These are the hydrocarbon constituents that are liquified and removed from the gas stream by the processing plant.

It is not the purpose of this paper to deal at length with the

chemistry and physics involved in the operation of a natural gas processing plant. Rather, two postulates will be provided to enable the non-technical reader to understand any theoretical discussions presented. These postulates are easily verified by an examination of certain basic engineering texts devoted to gas processing.[2] They are presented below.

Postulate 1. The heavier a hydrocarbon, the higher the temperature and the lower the pressure that it can exist in the liquid state.

Postulate 2. For any mixture which is being cooled at a fixed pressure to separate its components, there will be, at any normal process temperature, a certain percentage of the heavier components in the liquid state, with the remaining percentage of these components in the gaseous state. If the temperature is reduced, the percentage of each of the heavier components in the liquid state increases, and the percentage of these components in the gaseous state correspondingly decreases. Note that the lightest components in the mixture could remain almost entirely in the gaseous state unless the temperatures were lowered below normal process temperatures. (To do so would defeat the purposes of the process.)

The reader should now recognize refrigeration as one of the most important methods of processing natural gas. By lowering the temperature sufficiently, any required amount of the heavier hydro-

carbon components present in the raw gas can be liquified and separated from the gas stream.

Some of the elements (helium, oxygen, nitrogen) listed in Table I are generally found only in trace quantities and as such are not significantly detrimental to the gas quality to warrant processing for removal. There are three broad classes of components that must be removed, however, for various reasons explained below.

<u>Acid gases</u>. (Hydrogen sulphide and carbon dioxide.)[3] These components were originally removed to meet quality specifications. They both react chemically to form highly corrosive acids which can destroy plant equipment, pipelines, and utility equipment. Second, H_2S is a highly lethal gas which can kill instantly. Third, the presence of these gases lowers the heating value of the gas. Finally, one of the combustion products of H_2S is sulphur dioxide (SO_2) which is characterized by its highly offensive smell.

Economic considerations are now important in hydrogen sulphide removal since the hydrogen sulphide can react with controlled amounts of oxygen to form a very pure form of sulphur. In some plants the revenue from the sulphur produced exceeds the revenue from all other sources combined.

The specifications imposed by the pipeline companies are quite strict regarding the presence of acid gases. Trans Canada Pipelines requires that its purchased gas have no more than one grain of hydrogen

sulphide per one hundred cubic feet and a total sulphur content of no more than twenty grains per one hundred cubic feet. (There are six hundred grains of hydrogen sulphide in one hundred cubic feet of gas containing one per cent hydrogen sulphide.) Carbon dioxide maximums are usually set at two per cent.

Water. There are three reasons that water vapor must be removed from the gas before it is sold. First, although the water might be in vapor form as it leaves the wellhead, the cooling of the gas from reservoir temperature to ground temperature causes liquification of the water vapor in the pipeline. This water can form pools in low lying sections of the pipeline and cause "slugging" or uneven flow of gas and water through the line.[4]

Second, liquid water helps to form hydrates, which are complicated compounds of hydrocarbons, water and other substances such as hydrogen sulphide or carbon dioxide. These can form granular solids under certain temperature and pressure conditions which can completely plug a pipeline.

Third, the gas purchaser is obviously not going to purchase fuel gas containing a large amount of water, even if it is in the vapor state. Trans Canada Pipelines requires that the gas contain no more than four pounds of water per million cubic feet of gas purchased, when measured at 14.65 psia. and 60°F.

Natural gas liquids (including condensate). The purchaser does not want too high a liquid hydrocarbon content since it can create some of the same problems that water creates in pipelines. However, the main reason for extraction of natural gas liquids is often economic, from the gas producer's point of view. These liquids can sell for prices quite comparable to crude oil prices and as such provide an important incremental revenue to the plant owner. Trans Canada Pipelines requires that the gas sold to them must have all the liquids removed that would condense at a temperature of 15°F, when at a pressure of 800 psig. Many gas producers remove liquids that condense at much lower temperatures for purely economic reasons.

The pipeline companies do not want all the natural gas liquids removed however, as this would result in a very low heating value for the gas purchased. (The heavier the hydrocarbon, the higher its heating value.) Therefore a minimum B.T.U. content is usually specified. Trans Canada Pipelines often requires that the natural gas purchased by them have a minimum B.T.U. content of 950 B.T.U. per cubic foot.

II. OPERATION

Figure 1 shows a simplified schematic diagram of the operation of a typical natural gas processing plant. This section explains the chemical processes involved in the removal of the three classes of components noted above.

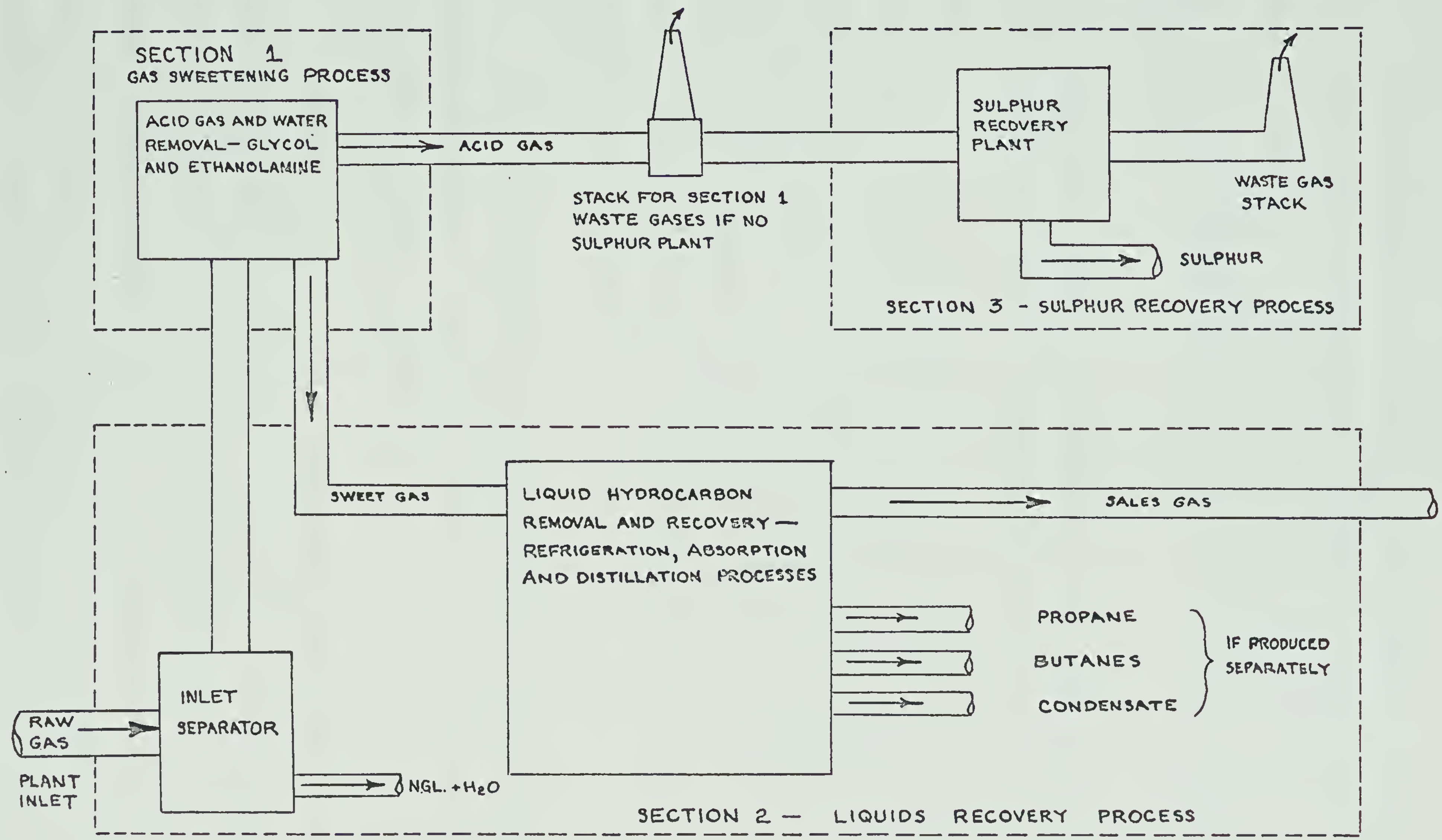

FIGURE 1 SIMPLIFIED SCHEMATIC DIAGRAM OF A CONVENTIONAL PROCESSING PLANT

It is necessary to stress that this study is concerned with the most common type of plant process employed in Alberta today. There are a few plants that employ different processing methods. However, the majority of plants use all or part of the process system described below.

The type of plant to be examined is one that takes the raw gas input and first removes any easily removable natural gas liquids and water in an inlet separator. This unit is essentially just a tank containing a series of baffles and screens. The rapid decrease in velocity causes any liquids entrained in the gas stream to fall out and attach themselves to the screens. They then drip down the screens and are drained from the tank bottoms. (Note that this process only removes entrained liquids; the heavier hydrocarbons still in the vapor phase are little affected.)

The inlet separator, as it is concerned with the separation of natural gas liquids from the gas stream, is included in Section 2 of the process diagram. (Figure 1.)

The raw gas, with some of its natural gas liquids removed by inlet separation, now moves back into Section 1 of the plant where it is treated for removal of acid gases and water. One of the most common processes used for this, and the only one discussed in this paper, is the glycol-water-ethanolamine (amine) process. Basically the glycol is used to absorb the water and the amine solution is used

to absorb the acid gas from the plant feed. It should be noted that water removal occurs after acid gas removal. (Dehydration costs are too small to be considered here.)

The basic amine process can use two types of solution--the monoethanolamine (M.E.A.) or the diethanolamine (D.E.A.) solution. The process equipment is basically the same for both solutions however, the plant using D.E.A. can utilize slightly smaller equipment in some cases.

The earlier plants were all constructed to use M.E.A. High amine losses occurred when the inlet gas contained carbon disulphide, carbonyl sulphide or a high percentage of the heavier hydrocarbons such as condensate. D.E.A. does not suffer from the same defects, although operating costs tend to be higher.

Opinion as to the overall superiority of either solution is divided, however, most processing engineers tend to favor D.E.A., especially where the conditions mentioned above are found to exist.[5]

Figure 2 shows in simplified form the operation of this section of the plant.

The inlet feed enters the contactor at the bottom and flows countercurrent to the flow of liquid amine solution. The trays noted on the diagram are "bubble trays" whose sole function is to provide effective contacting of the vapor and the liquid, thus increasing the efficiency of the operation.

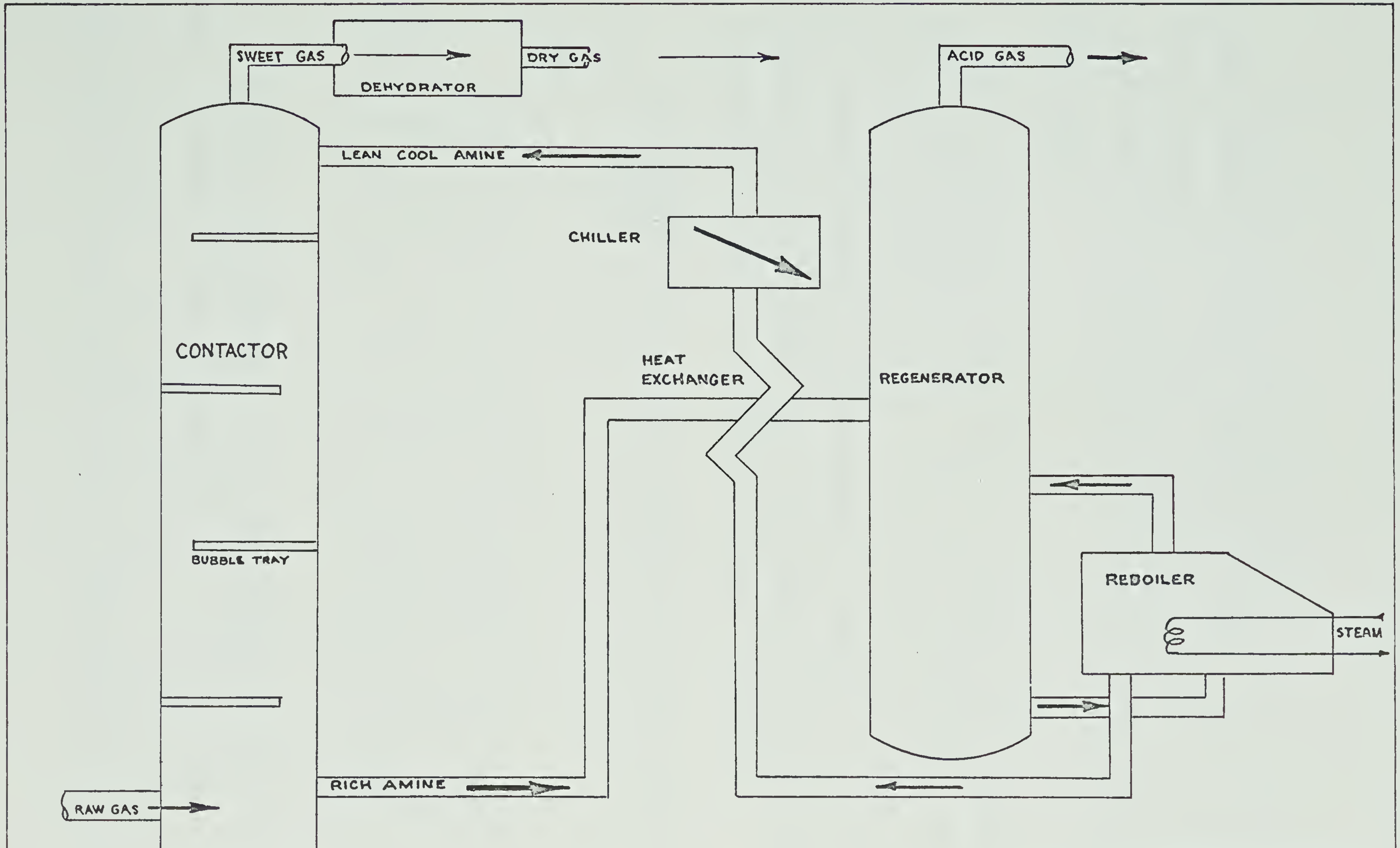

FIGURE 2 SIMPLIFIED SCHEMATIC DIAGRAM OF AMINE SECTION (SECTION 1)

The gas leaving the glycol-amine section is essentially free of carbon dioxide and hydrogen sulphide and the water content is reduced to below specification maximums. (The hydrogen sulphide content usually ranges from .050 grains to .25 grains per one hundred cubic feet of gas.)

The rich amine solution leaving the bottom of the contactor is sent to the regeneration tower. Basically, all that happens here is that heat is applied to the rich solution, driving off the acid gas and leaving the amine in a "lean" condition, ready to be circulated through the contactor again. Note that cooling of this amine stream is necessary to increase the acid gas absorption capability.

The acid gas, now completely separated from other elements, is flared to atmosphere or sent to the sulphur recovery section if present.

Process Section 2 in Figure 1 is the section which removes natural gas liquids from the sweet gas leaving the amine section. There are several ways to do this, but this paper is concerned only with those plants which use refrigeration or absorption processes.

If the gas is chilled below its natural temperatures it is easy to see that the heavier components will liquify and settle out of the gas mixture. (See Postulates 1 and 2.) The colder the natural gas becomes, the more liquids produced. Further, any given percentage

of the heavier hydrocarbons will always condense before the same percentage of the lighter ones.

In order to remove a high percentage of the heavier hydrocarbons such as condensate, butanes and propane, low temperatures and hence large refrigeration units would be required. However, at the temperature required to condense 95% of the propane, for example, a significant percentage of the methane is also condensed. Thus a simple refrigeration (or "shallow cut") process is not sufficiently precise for use in high recovery schemes, and is used only to recover condensate along with small amounts of propane and butanes. In order to recover the majority of propane and butanes, a more specific process is required.

For a "deep cut" process, or one in which the majority of the natural gas liquids are removed, an absorption process is used in conjunction with the refrigeration process. The result is the conventional cold oil absorption process. This process is very similar in operation to the amine section process described earlier.

Figure 3 shows a simplified model of a cold oil absorption process. The rich natural gas enters the absorber in countercurrent flow to the lean cold oil, which absorbs its heavier components. The gas stream then leaves the plant, having been completely processed to sales gas specifications. The oil stream, now loaded with natural gas liquids, is fed to the recovery section where it is separated by a

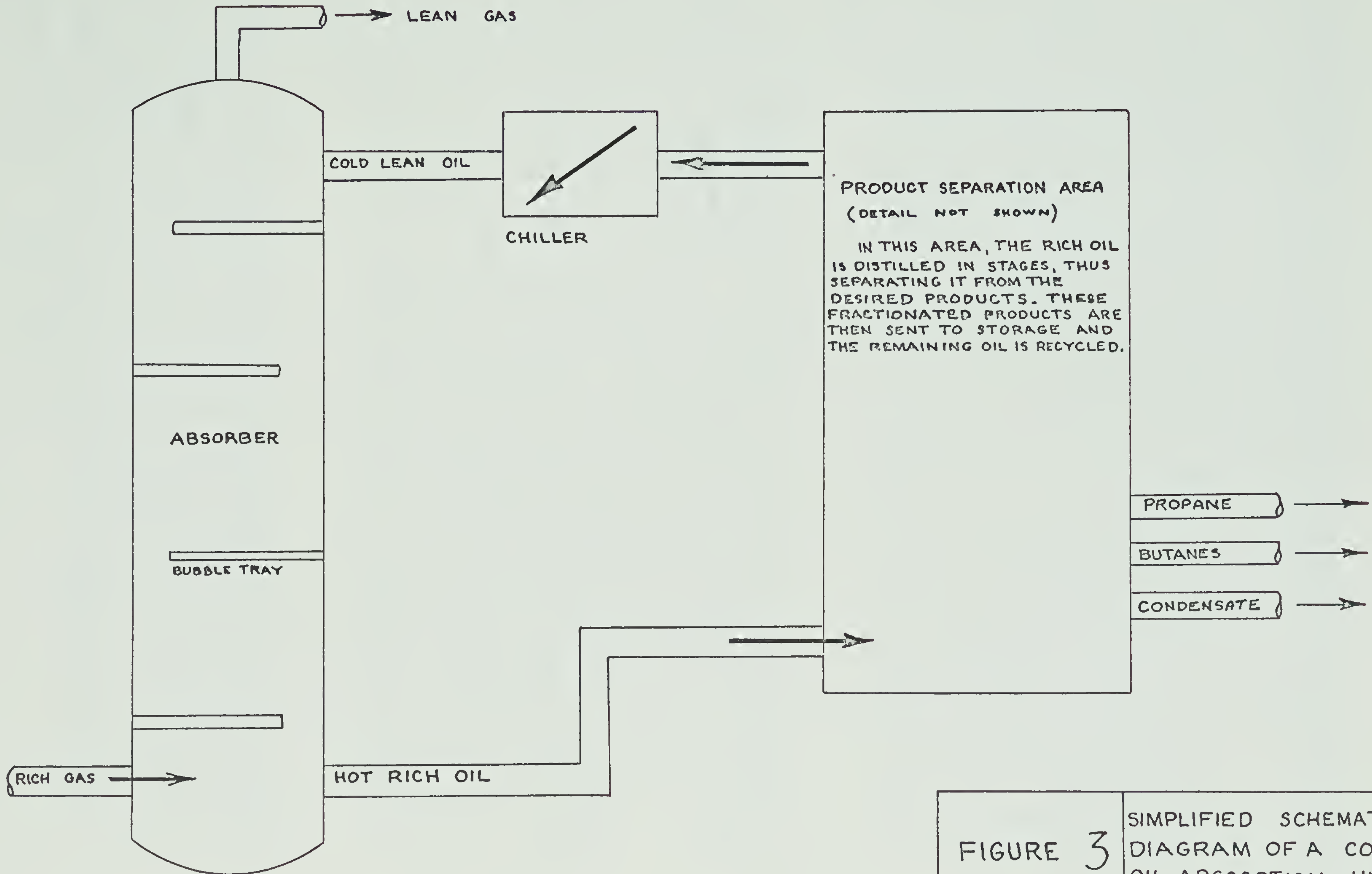

FIGURE 3 SIMPLIFIED SCHEMATIC DIAGRAM OF A COLD OIL ABSORPTION UNIT

distillation process into lean oil and the various natural gas liquids. The oil is then chilled and recirculated through the system.

Section 3 of Figure 1 shows the optional sulphur recovery unit. If economics or air pollution regulations dictate that sulphur should be produced from the acid gas stream, this type of unit is constructed to use the amine plant acid gas as feed. There are three basic types of process used for this: (1) "In Line" process (Claus), (2) "Bypass" process (Modified Claus), and (3) "Jefferson Lake" process (Recycle Process).

The "Jefferson Lake" process is not common and is therefore not considered in this thesis. The mechanics of the other two processes both involve the burning of H_2S with controlled amounts of air, followed by catalytic conversion. The "Bypass" process is generally used for low concentrations of H_2S in the acid gas stream (below 50%) while the "In Line" process is used for higher H_2S concentrations. In addition, the "In Line" process is rarely used where the acid gas flow rate is less than 50-100 MMcf/Day, because of combustion efficiency problems.[6]

B.S. & B. sulphur plants are modifications of conventional "In Line" or "Bypass" plants. The B.S. & B. process incorporates heat exchange equipment between the sulphur plant and the amine regeneration system, thus eliminating the need for a separate amine regenerator heat source. (See Figure 2.)

If a sulphur recovery section is not included in the plant, the highly poisonous acid gases must be incinerated with natural gas and disseminated through the use of a large stack. These stacks range up to three hundred feet in height in order to reduce pollution problems in the surrounding areas. Fortunately, if the acid gas contains even as little as 5% H_2S, sulphur recovery can be economically feasible, which reduces the potential pollution problem considerably, hence diminishing the required stack height.

The reader can possibly see several different combinations of process sections which could be used for gas processing plants. For example, a plant which was treating a "sweet" gas would not require any acid gas treating facilities, although water might still be a problem. When for economic or other reasons, the quantity of natural gas liquids to be removed is small, a simple refrigeration plant may be all that is required. When large quantities of natural gas liquids are present, and markets are available, the more complex "deep cut" process would probably be in order.

Several designers have divided Section 2 (liquids recovery section) of Figure 1 into two sections--gas processing and liquids processing.[7] Although a valid case can be made for this arbitrary decision, this writer found that the use of one section instead of two is adequate for the purposes of this thesis.

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[1]Other elements can exist in "trace" quantities.

[2]Petroleum Industry Training Service, op. cit., Chapter 1.

[3]There are sometimes other "trace" sulphur compounds present, such as carbonyl sulphide and carbon disulphide.

[4]For this reason, small dehydrators are often installed right at the wellhead.

[5]Petroleum Industry Training Service, op. cit. p. 12-2; James W. Estep and Edward W. Plum, "Economics of Sour Gas Industry" (Paper read at the fall meeting of the Society of Mining Engineers, Los Vegas, Nevada, October 1967) pp. 9-13.

[6]Private communication with James W. Estep, Texas Gulf Sulphur Company, Calgary. (See above.)

[7]This division was found in several processing plant proposals examined by the writer.

CHAPTER III

CAPITAL COST ESTIMATING PROCEDURES USED TODAY

This section investigates the present methods of capital cost estimating used for natural gas processing plants in Alberta. These range from detailed estimates based on detailed final designs to the educated guesses of process engineers. This paper deals with the detailed estimate first since this will enable the reader to understand some of the assumptions made in the more abbreviated estimating methods discussed later.

I. THE DETAILED COST ESTIMATE

The detailed cost estimate is a by-product of the processing plant design. Therefore, a brief outline of the plant design procedure is necessary at this point.

Gas plants are not often built to process the gas from only one well. Generally speaking, it would not be economically feasible, and even if it was, very few managements would authorize a gas plant for only one well if the total field potential had not been explored. Therefore, the gas plant is not contemplated until at least part of the field has been delineated.

Assuming that the gas reserves are established and the total

daily production of the field has been calculated, the preliminary design of the proposed processing plant is commenced. This is usually done within the company intending to produce the gas, primarily for a rough approximation of the economic feasibility of the total gas production project. If the project passes this first hurdle, then the detailed design begins. This can also be done within the company; however, most companies contract out the detailed design to consultants or processing plant manufacturing specialists.

The input data for a processing plant design can be divided into three broad classes of information: (1) Inlet gas Parameters, (2) Output Gas & Products Specifications, and (3) Environmental Data.

The inlet gas parameters and the output gas and products specifications are of course interrelated in that the composition of the inlet gas dictates to some degree what can be recovered. Probably the most important variables however, are the design flow rate and the composition of the inlet gas, given the quality specifications required by the purchasers.

Inlet Gas Parameters

Design flow rate. The maximum allowable flow rate of any Alberta gas well is restricted by government regulations; however, the effective maximum used is set by the pipeline companies' stipulated maximum daily volume. This volume, for the total field, generally sets the design "Q" or flow rate required for the proposed gas plant. It

should be noted that often excess flow capacity is built into the plant in anticipation of future field development.

Pressure and temperature of the inlet gas. Pressure considerations dictate the size and strength of the piping and vessels used in the plant and therefore can play an important role in the design and cost of a processing plant. In the majority of plants studied however, the pressures were found to be within a fairly narrow range and the standardization of equipment sizes tended to negate pressure as a significant variable although it must, in theory, be considered. Note that "capital cost", as defined in this study, does not include the cost of the compression facilities sometimes required to compress the gas to pipeline pressures.

Temperature of the inlet gas is significant because the gas must be cooled down to the specified sales gas dewpoint or lower in order to recover the liquid products existing in the gas as vapors. Therefore, the higher the inlet gas temperature, the more refrigeration equipment required. It is interesting to note that because of the length of gathering system required, the inlet gas enters the plant at essentially ground temperature. Therefore some provision must be made for summer-winter ground temperature fluctuations when designing the plant.

Inlet gas analysis. This is a very significant part of any processing plant design. Table I notes the constituents possible in the plant inlet gas and it is essential that an accurate analysis be

done by a chemical laboratory to enable the process engineers to design a process to fit the particular inlet gas composition. For the liquids such as propane (C_3), butane (C_4), and pentanes plus (C_{5+}), this can be especially important since all liquids processing facilities are sized for specific flow rates. The percentages of each component in the inlet gas stream are usually converted to "barrels of liquid per million cubic feet of gas" (BBL/MMcf) for calculation purposes.

The water content of the gas is usually excluded from the gas analysis and material balance calculations. However, the water content of the gas is required data for design because most of it must be removed before sale.

Acid gas content is determined and if above the allowable maximums, a gas sweetening section must be planned for the plant. It should be noted that there can be some relaxation of the requirements as far as carbon dioxide is concerned, by negotiations with the pipeline companies. However, the hydrogen sulphide requirements are very rigid.

Gas Outlet and Product Requirements

The gas transmission companies who purchase the gas have set down specific quality specifications. In summary, these are:

TABLE II

TYPICAL GAS QUALITY SPECIFICATIONS[1]

Item		Typical Specifications
Heating Value	(Min.)	950 BTU/Lb.
Hydrocarbon Dewpoint	(Max.)	$15^{o}F$ @ 800 psig.
Water Content	(Max.)	4 lb./MMcf
Sulphur (& Compounds) Content	(Max.)	1 grain H_2S, 20 grains total sulphur per 100 cubic feet
Carbon Dioxide	(Max.)	2%
Temperature	(Max.)	$120^{o}F$
Delivery Pressure	(Min.)	900 psig.

These specifications provide an operating envelope for the proposed plant. The actual hydrocarbon dewpoint is quite possibly lower, however, since the natural gas liquids removed are quite valuable in themselves. (Note that the BTU minimums could limit the maximum liquids recovery.) An economic study is therefore required to find the optimum size of liquid recovery plant given the input data and the forecasted prices of the liquids recovered. In practice, this is only approximated and in some cases, the decision to recover a certain quantity of liquids from the gas is made on an entirely arbitrary basis, or is based perhaps on current industry standards.

Once the required liquid output of the plant has been specified the process engineers can design this section, often with the aid of computer programs. Note that it is quite possible for a plant to be designed with no natural gas liquids recovery. This could occur if the percentage of natural gas liquids present in the inlet stream is very low.

If the inlet gas is sour and appreciable quantities of hydrogen sulphide are present, two related studies must be made. First, the sour gas must be removed. The most common technique in Alberta is that of the "amine" process. This process, discussed in Chapter II, is relatively simple and the sizing of this section is, as expected, directly related to the amount of amine solution circulated, which is itself directly related to the amount of acid gas processed. Note that a very common process adds a glycol dehydration unit to remove the water from the sweetened gas.

Hydrogen sulphide can be processed to produce pure sulphur and if sufficient quantities are producible, the sulphur plant is included as part of the design work. Manufacturers of sulphur plants have designed these plants on a package basis, and the major information required here is acid gas production and percentage hydrogen sulphide. This is becoming a specialized field and most oil companies are quite prepared to subcontract this section of the plant to a specialist. Pricing information on these plants is quite good and most manufacturers

can give an accurate cost estimate on relatively short notice.

Environmental Data

The environmental data includes material and labor costs, climatalogical factors and geographical factors such as proximity to transportation facilities and industrial centres. General economic factors such as inflation or recession can also have a significant effect on the timing of construction.

Once all the above information has been assembled and processed, the sizing of all major equipment items can be completed and piping requirements can be established. Power and heating requirements are then determined and compressors, pumps, boilers and motors are selected, usually from standard sizes available by manufacturers.

The selected site is examined and soil surveys are conducted. The plant physical layout is selected and foundations and buildings are provided where required. Control, instrumentation and safety equipment are designed in detail at this stage.

When the plant has been completely designed and detailed engineering drawings and specifications are available, the cost estimating procedure used is relatively simple but quite tedious. Each item is individually priced, from major equipment down to piping, valves and wiring. Man hours required for assembly are estimated from past experience. Construction expenses, such as camp costs, power

costs, sales taxes on purchased equipment, and supervision costs are added. When the plant is built under contract (as most are), the contractors' overhead, contingency factors, and profit are included in the total cost estimate.

It should be noted that the cost estimate arrived at using this method is usually within plus or minus five per cent of the actual completed cost of the processing plant.[2]

II. AREA FACTOR METHOD

The area factor method is a simplification of the detailed cost estimate. It involves the actual pricing of the major equipment (as delivered), and the application of a multiplication factor to arrive at the total installed plant cost. This factor, when applied to the total plant, generally ranges between 3.5 and 4.5, depending upon the type of plant and environmental factors. Writing in 1948, Lang proposed an area factor of 4.74 for estimating the capital costs of liquid process chemical plants.[3]

A refinement of this method uses one "area factor" for each type of major equipment. The following factors are representative of those used in Alberta gas processing plants:[4]

TABLE III

AREA FACTORS FOR ALBERTA GAS PROCESSING PLANTS

Item	Area Factor
Fractionating columns	4
Pressure vessels	4
Heat exchangers	3.5
Fired heaters	2
Pumps	4
Compressors	2.5
Instruments	4
Miscellaneous	2.5

It should be noted that these are only approximate factors which are subject to large variations because of other variables such as automation and instrumentation required. The accuracy of estimates based on this method is in the order of plus or minus thirty per cent.[5]

III. THE "POINT SIX" RULE

The "point six" rule states that the ratio of the costs of two projects will approximate the ratio of their sizes, raised to the six-tenths power. Mathematically, this is stated as:

$$\frac{\text{cost of A}}{\text{cost of B}} = \left(\frac{\text{size of A}}{\text{size of B}}\right)^{0.6}$$

This is a common rule of thumb in the industry. It could be used effectively where "size" is a specific defined function of project capacity, and where the type of construction, operating variables, and materials used for construction are similar for the plants being compared.[6] Unfortunately, the accuracy of this method is not satisfactory when applied to natural gas processing plants. Errors of thirty to one hundred per cent have been observed by the writer.

One of the major reasons for the poor performance of this rule is that the proper definition of size or capacity is difficult if not impossible to formulate when so many variables influence the size and cost of a plant. Size would have to be a specific function of all significant variables influencing the total plant cost and, to date, this function has not been derived. For most calculations using this rule, the flow rate (Q) is used as the only variable differentiating roughly similar plants, and an adjustment is made to consider any other factors the estimator believes to be pertinent. It is not surprising that this rule rarely provides good estimates.

Chilton suggests a factor of .33 instead of .6 when dealing with the removal of H_2S from natural gas.[7] His unit of capacity or size function is the natural gas charge. Note that this does not consider

the percentage of H_2S in this charge.

The results of this study indicate that the six-tenths rule may not be correct even when a satisfactory size function is present. The resulting curves do not follow the exponential pattern, even when the plant is divided into functional sections.

Max Peters supports the above statement in his discussion on chemical plant capital costs.[8] He suggests that a seven-tenths factor replace the six-tenths factor for total plant estimates but warns that the factor does not remain constant, but increases from approximately three-tenths for very small plants to one for large plants utilizing multiple units.

There is a definite use for the "point six" rule, however. When sizing individual pieces of equipment, this rule can be reasonably accurate. The exponent is not usually .6, as the list of equipment exponents shown in Table IV indicates.[9]

It should be noted that these exponents have been derived statistically over the whole range of refinery construction in Canada and the United States. Therefore, the use of these exponents for gas processing plants in Alberta could result in large errors.

TABLE IV

TYPICAL EXPONENTS FOR INDIVIDUAL EQUIPMENT TYPES

Equipment	Exponent
Reciprocating Compressors	.96
Fintube Heat Exchangers	.58
Kettle Reboilers	.75
Pressure Vessels	.68
Refrigeration Units	.72
Horizontal Tanks	.67
Process Towers	.79

Another caution should be noted here. This rule is not to be used over too large a range. The very small equipment can be at the minimum cost point at which any further reduction in size would not decrease the cost. Some evidence of this phenomenon is provided in the next chapter. Second, the very large equipment might be designed with two or more parallel trains instead of one large unit. (This is commonly done to provide flexibility and reduce losses during breakdowns.) This situation would not be adequately estimated by the six-tenths rule.

IV. GRAPHICAL METHODS

The estimating methods based on cost charts are not too numerous and for the most part, have proven unsatisfactory for estimating purposes in Alberta. Three of these are mentioned briefly.

J.M. Hillsman has derived a series of curves relating capital costs to certain operating parameters of gas plants.[10] There are three major problems associated with the use of this study. First, the capital costs all pertain to plants constructed in the southern United States and thus would not necessarily reflect capital costs in Alberta. Second, the study deals only with cold oil absorption units, and third, the parameters required to use the curves require considerable prior calculations thus making this method quite time consuming. The author estimates the accuracy of this method at plus or minus ten per cent.

Gino Giusti, in his article on sulphur recovery plants, provides some cost curves on amine sections.[11] Although these curves are not supported in any way by his article, and seem to be based on the "point six" rule, one of his basic assumptions is of interest and therefore his curves are provided for comparison and discussion purposes in Chapter IV.

Estep has provided capital cost curves relating to sour gas plants.[12] His data was of limited use for the purposes of this thesis

as all his studies were done at a fixed plant inlet flow rate and there was no provision made for the recovery of natural gas liquids. However, his discussions on the amine and sulphur recovery sections were most beneficial to this writer.

V. THE EDUCATED GUESS IN COST ESTIMATING

This "method" is probably the most commonly used for preliminary estimates and the method most susceptable to large error. The writer has observed errors up to three hundred per cent. There is very little to discuss about this. The ability to estimate comes with experience and even this is not enough to provide consistently realistic estimates. It was to replace this "method" that the study contained herein was first proposed. The next chapter investigates the problems involved in rapid estimating and derives a method of quick cost estimating that improves on other rapid methods now in use.

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[1] These specifications were taken from the most common Trans Canada Pipelines contract in force at the time.

[2] Petroleum Industry Training Service, op. cit. p. 20-1.

[3] Hans J. Lang, "Simplified Approach to Preliminary Cost Estimation", Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 12-14.

[4] Petroleum Industry Training Service, op. cit. p. 20-6.

[5] Petroleum Industry Training Service, op. cit. p. 20-1.

[6]Max S. Peters, Plant Design and Economics for Chemical Engineers (McGraw-Hill Series in Chemical Engineering, McGraw-Hill Book Company, Inc, 1958) pp. 94-96.

[7]Cecil H. Chilton, "Six-Tenths Factor Applies to Complete Plant Costs", Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 282-284.

[8]Max S. Peters, op. cit. pp. 100-102.

[9]Petroleum Industry Training Service, op. cit. p. 20-3. Often weight is used as the parameter of size for this calculation.

[10]J.M. Hillsman, "Faster Cost Estimates for Gas Plants", Hydrocarbon Processing & Petroleum Refiner, Vol. 42, No. 4, April 1963, pp. 171-174.

[11]Gino P. Giusti, "Sulphur Recovery Processes", Oil and Gas Journal, February 22, 1965, p.99.

[12]James W. Estep and Edward W. Plum, "Economics of Sour Gas Industry". (Paper read at the fall meeting of the Society of Mining Engineers, Los Vegas, Nevada, October 1967.)

CHAPTER IV

A RAPID METHOD OF CAPITAL COST ESTIMATING FOR GAS PROCESSING PLANTS

I. INTRODUCTION

The object of this study was to replace all cost estimating procedures, except the detailed cost estimate, with a method that is very rapid and considerably more accurate than the methods now used. Of prime concern was the initial cost estimate used for economic evaluations. These evaluations are sometimes required within days, with some managements being prepared to commit company funds on the basis of such evaluations. Obviously, there is no time for a total plant design; however, a reasonably accurate cost estimate must be provided for financial decision making purposes. The original target accuracy (the estimate compared with the actual cost) was set at plus or minus twenty per cent which compared favorably with the error range of thirty to three hundred per cent found by the writer.

II. LIMITS TO THIS STUDY

Some limits had to be set upon the scope of this study. The following limits were present from the outset, others were required as the study progressed.

The processing plants studied are all situated in Alberta. This limitation was imposed primarily because of the lack of infor-

mation available to the writer on out of province plants. However, it is suggested that the results of this study could also be applied to British Columbia and Saskatchewan processing plants if appropriate corrections were to be applied. For example, transportation costs for equipment used in constructing a gas plant located in extreme northern British Columbia would probably be an important factor in the overall cost.

None of the plants studied are isolated in the sense that transportation to or from the plant is difficult. This is to be expected to some degree since the processing plant is not normally constructed until the gas field is at least partially developed and this in itself tends to develop the area and its transportation facilities.

There are several different types of processes used to convert the raw gas into pipeline specification gas. From the outset, it was obvious that there was a preponderance of one general type. If the gas was sour, it was sweetened by contacting it with an amine solution which effectively removed one hundred per cent of the acid gas components. A glycol unit subsequently dehydrated the sweetened gas. The sulphur plants all used the acid gas stream as feed and burned these gases to produce elemental sulphur. The sweet gases leaving the amine section were treated by refrigeration and sometimes absorption in order to remove the natural gas liquids.

Other types of processes, such as the dry dessicant sweetening

and water removal processes, or the "molecular sieve" sulphur removal process, fall outside the scope of this study. This did not materially affect the results however, as there are very few plants designed around these processes in Alberta today.

Plant size was another significant limitation which was imposed to keep the study within manageable proportions. During the initial phases of the study a list of Alberta gas plants was examined and it was observed that a very high percentage of these plants had inlet flow rates of less than one hundred million cubic feet per day (100 MMcf/D). Therefore, this was the upper limit on size originally chosen for this study. As previously discussed, inlet flow is only <u>one</u> measure of size and the limitations of choosing this measure were clearly shown in later calculations.

Some gas processing plants actually process a two phase flow of oil and gas. This was thought to be a potential problem but the particular two phase plant included (Ferrier), proved to be no real problem since the liquid was separated at the inlet separators (as is normal practice) and the rest of the plant is concerned only with gas processing. The Ferrier plant did not deviate from the general function derived for all gas plants, which supports the above conclusions.

Crude oil, when it is gathered into tank batteries, has "in solution" a rich natural gas which evaporates at atmospheric pressure. This "solution gas" is often collected, compressed and processed, just

as ordinary gas is processed. Plants processing this gas were included in this study. Note however, that the capital cost of the required compression facility is outside the scope of this study.

The final limitation was placed on the study because of lack of satisfactory information available to the writer. The company employing the writer provided all available files on owned or partially owned plants, but the simple fact was that even these were not complete. Personnel of other companies were most reluctant to provide information and little co-operation was forthcoming.

III. DATA COLLECTION

Appendix A summarizes the data collected on eleven gas processing plants in Alberta. Of these, ten were selected for further study. The Carson Creek processing plant was suspect from the beginning because of its radically different operating pressure, and therefore it was not included in the original study.

The information included in the data summaries constituted all the factors that were thought to affect the capital costs of the plants studied. It was recognized early in the information search process that many of the items tabulated would have little effect on this study. First, a parameter having a value common to all the plants studied could not be expected to explain differences in total cost. Second, there were some parameters that were not believed to be important to the capital costs of the plants studied. However, all available data

was recorded as the writer was uncertain just which variables would in fact be significant.

The liquids portion of the gas analysis was converted to a common denominator for analysis. This was the liquid to gas ratio "barrels of liquid per million cubic feet of sales or residue gas". The outlet quantity was preferred over the inlet quantity because of the large variance expected in the acid gas content of the inlet gas. If, for example, the acid gas component made up fifty per cent of the inlet gas, the liquid to inlet gas ratio would not be truly indicative of that entering the liquid recovery section of the plant. The liquid to outlet gas ratio, although still not exact, provided a much more satisfactory parameter.

The data on the actual plant costs was drawn from various sources. Cost breakdowns were available on very few of the plants and often considerable research was required to break down the available information for use in this study.

One of the problems encountered was that several companies provided the total cost for the processing plant, gas gathering system, the loading and transportation facilities for marketing of liquid products, wellhead equipment and, in sum, all capital costs pertaining to producing the gas field. The only costs that this study was concerned with were those of the processing plant itself, as defined in Chapter I.

Total plant capital costs (as previously defined) were successfully determined from the information available. Unfortunately, this was not the case when a cost breakdown within the plant was attempted. There was, generally speaking, no available detailed cost breakdown that could be used to divide the total costs into the cost of each section (as defined in Chapter II). The one exception to this was Imperial Oil's "Boundary Lake" processing plant, for which the contractor (Stearns Rogers of Canada, Ltd.) provided a detailed cost breakdown for every item provided in the plant. Although the allocations of some costs were not acceptable for the purposes of this study, this breakdown was most useful.

IV. INITIAL DATA PROCESSING

The first step required was to convert all costs into common units of purchasing power. A price index was required that would specifically indicate price changes for gas processing plant construction in Alberta. Two published indices were first examined in this regard.

The "Engineering News Record" cost construction index combines the price indices for structural steel, concrete, lumber and labor into one composite construction cost index.[1] Unfortunately it was found to be too general for the purposes of this study as it applies to all types of construction and is based on the prices present in twenty American cities. However, this index could possibly indicate

a trend in price changes. The converted index is shown on Table V and graphically on Figure 4.

The "Nelson Refinery Construction Cost Index" was thought to provide a much more suitable measure of price changes for Alberta gas processing plants.[2] It is divided into three segments: (1) "miscellaneous equipment" such as pumps, compressors, electrical machines, engines, heat exchangers, and instruments, (2) "major materials", and (3) "labor". Table V and Figure 4 include this index.

The two published indices discussed above reflect price changes in various sectors of the American economy. If one were interested in the price changes occurring in the Canadian construction industry, the construction index provided by the Dominion Bureau of Statistics would probably be more suitable. However, there is no Canadian index pertaining directly to the refining industry, and for this reason, the "Nelson" index is considered to be the most appropriate published index for use with gas processing plant capital costs. However, it is to be remembered that it does not pertain to Canadian prices, let alone prices within a specific geographical area such as Alberta or within a specific segment of the refining industry.

The validity of these and other published indices was discussed with several consulting engineers and processing plant contractors in Calgary. The general conclusion was that although the "Nelson" index was perhaps the most valid of the published indices, <u>no</u> published index

TABLE V

COST INDICES CONSIDERED FOR THIS STUDY
1958 TO 1967
BASE YEAR 1967*

	Engineering Digest	Nelson Refinery	Pacific Petroleums	"SELECTED" INDEX	
				Index No.	Conversion Factor
1967	1.000	1.000	1.00	1.00	1.00
1966	.950	.925	.90	.90	1.11
1965	.905	.885	.80	0.81	1.23
1964	.870	.855	.71	0.73	1.37
1963	.835	.825	.63	0.67	1.49
1962	.810	.805	.60	0.65	1.54
1961	.790	.790	.58	0.63	1.59
1960	.765	.775	.56	0.61	1.64
1959	.740	.750	.55	0.59	1.70
1958	.705	.725	.50	0.55	1.82

*Rounded to nearest .005.

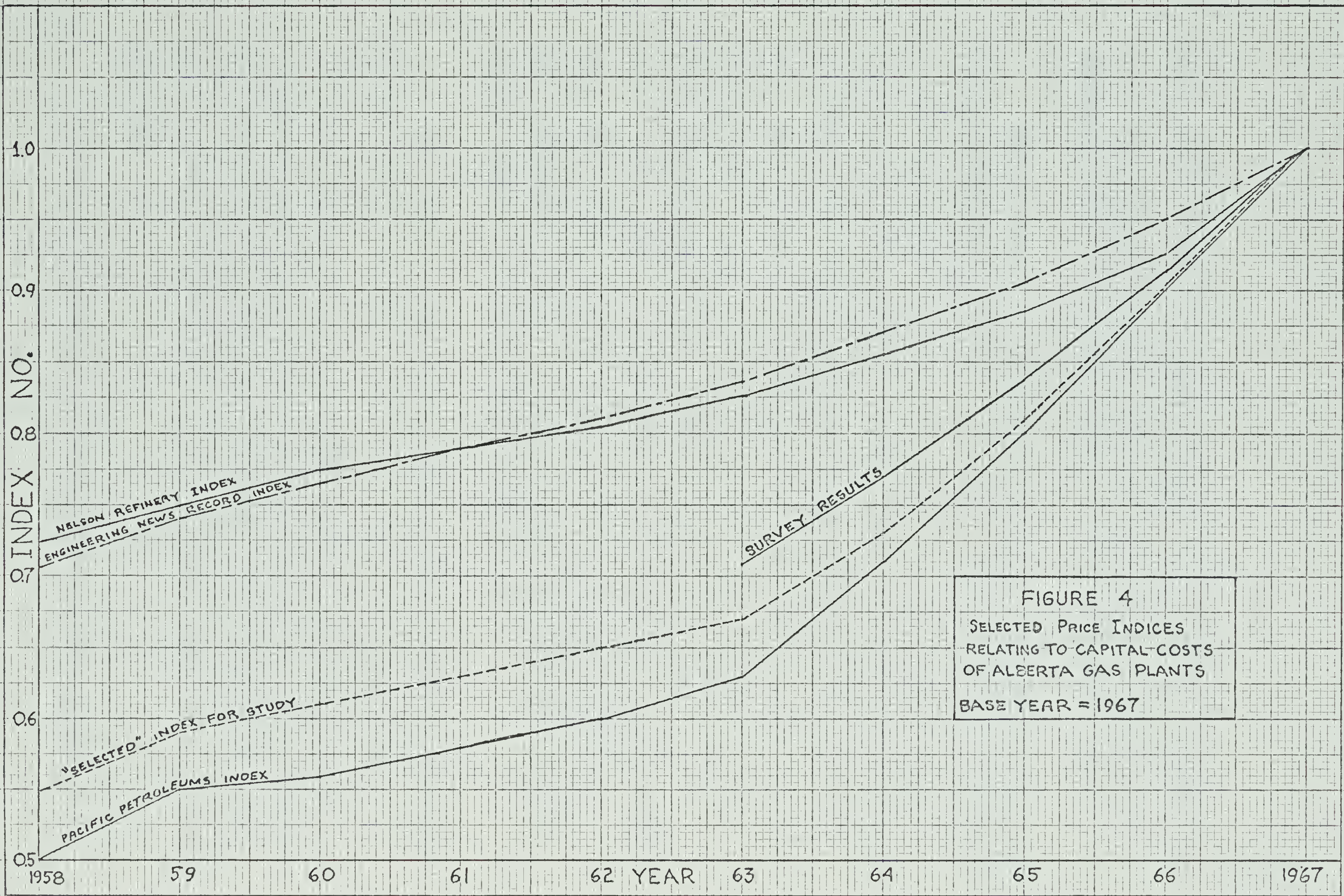

FIGURE 4
SELECTED PRICE INDICES
RELATING TO CAPITAL COSTS
OF ALBERTA GAS PLANTS

BASE YEAR = 1967

would properly reflect the price changes in the Alberta gas processing industry. Therefore, further examination of published indices was abandoned in favor of the derivation of a more appropriate index.

Capital costs in the Alberta natural gas processing industry are heavily influenced by demand. There are a limited number of contractors in Alberta available for processing plant construction. When these firms are fully committed to various projects, any additional projects can become very expensive because of large overhead increases and because there is little incentive for the contractor to accept the additional project except with a very large profit. On the other hand, there have been cases where contracting firms have tendered a "no profit" bid in order to retain their engineers and pay their overhead costs during slack construction periods. It has been suggested by some contractors that the level of construction activity could vary the bid prices by as much as thirty-five per cent.

The level of construction activity in other process industries such as the pulp and paper industry or the fertilizer industry also affects the prices in the gas processing industry. This is to be expected as some gas processing plant contractors are active in these other fields and the demand for their services in these other fields would influence the time and resources available for the gas processing industry.

Generally speaking, the Alberta gas processing industry has

grown rapidly during the last four or five years. Noting that there has been no large influx of contractors into this field, it is only natural to expect a period of rapidly increasing prices. This increase would not be shown by any of the more general published indices.

The first index felt to be representative of price changes in the Alberta gas processing industry is also shown on Figure 4 and Table V. It was derived from discussions with personnel at Pacific Petroleums Ltd. who were in a position to observe the price changes over the past ten years. Although the data was not as objective as the other two indices, it had the one advantage of being applicable solely to Alberta.

The main conclusion reached during discussions of Alberta processing plant price changes was that prices had doubled over the past ten years, and that there was a relatively flat increase during the period between 1960 and 1963, followed by a period of rapid price changes that had not abated at the end of 1967. Note that this trend is confirmed by the other indices provided.

The major difference between the "Pacific Petroleums" index and the published indices plotted on Figure 4 lies in the magnitude of the estimated price changes during the last four years. Note that the slopes of all three indices are quite similar prior to 1964. Therefore, the only serious problem to resolve was that of selecting a satisfactory index to measure price changes during the period 1964-

1967 inclusive.

Initially a compromise was made between the two published indices and the "Pacific Petroleums" index as it was felt that the most satisfactory curve probably lay between these two extremes. More weight was given to the "Pacific Petroleums" index after consideration of the factors discussed previously. The "selected" index is shown in column four of Table V and as the dashed line in Figure 4. This was the index finally chosen for use in this study.

Subsequent to the initial completion of the study, it was thought advisable to check the validity of the "selected" index. Unfortunately, the lack of data precluded any objective verification. However, the writer felt that a reasonably broad based survey of the opinions of consulting engineers, contractors and oil company executives could be used to confirm or deny the validity of the "selected" index. The results of this survey are shown in Appendix B and plotted on Figure 4.

The survey was directed only at the problem of verifying the rapid price increases thought to exist during the past four years as there was little conflict in the indices prior to this period. The reader will note that the maximum difference between the "selected" index values and the values obtained in the survey was 3.9% in 1963. Considering this small difference and the obvious approximations involved in all these index estimates, it was felt that the original

"selected" index was satisfactory for the purposes of this study.

The appropriate conversion factors for the "selected" index are also provided in Table V. These conversion factors are used to convert all costs into 1967 costs and are simply the reciprocals of the index numbers.

Table VI shows the adjustments made to the original cost of the ten plants studied and the total adjusted costs on which this study is based.

V. ANALYSIS

From experience, it was obvious that the first operation required to be performed on the data was the separation of plants requiring no amine section (sweet gas plants) from those requiring this section (sour gas plants). The amine section is quite separate from the rest of the plant, and is quite costly in itself. The five sweet gas plants were selected for initial study since the liquid recovery process is common to all plants and these plants had no other costs to complicate the analysis.

Sweet Gas Plants

A cursory examination of the data indicated no apparent reason for the presence of an amine section at the Boundary Lake plant. The carbon dioxide content was well below pipeline specifications and the

TABLE VI

ADJUSTED PROCESSING PLANT COSTS
(All Costs in Thousands)

Plant	Construction Date	Unadjusted Cost	Conversion Factor	Adjusted Cost	Total Adjusted Cost
Ferrier	1964	414	1.37	571	571
Kaybob	1962	400	1.54	616	
	1966	520	1.11	577	1193
Minnehik-Buck Lake	1961	1170	1.59	1860	
	1965	681	1.23	838	
	1967	445	1.00	445	3143
Paddle River	1966	1546	1.11	1716	1716
Ghost Pine	1967	1965	1.00		1965
Retlaw	1964	244	1.37	334	334
Gilby	1959	394	1.70	670	
	1965	754	1.23	927	
	1966	18	1.11	20	1617
Sylvan Lake	1963	890	1.49	1328	
	1967	1100*	1.00	1100	2428
Lone Pine	1967	3160	1.00		3160
Boundary Lake	1964	1525	1.37		2089

*The Sylvan Lake plant expansion cost was based on a "turnkey" contract bid of one million, fifteen thousand dollars accepted by the producers. Although this plant is not constructed yet, the bid was made in June 1967 and is based on 1967 prices. The sum of eighty-five thousand dollars was added as a contingency factor to cover non-contracted costs and this would definitely be subject to price level changes. However, the resulting error would be small and was therefore ignored in cost adjustments.

hydrogen sulphide content was barely measurable. Further investigation revealed that the gas was being used as fuel for a gas turbine power station and that due to the corrosion problem at the turbine operating temperatures, no hydrogen sulphide was allowed. This fact would allow one to obtain an excellent minimum amine section cost. (This aspect is discussed later.)

Although it was obvious to the writer that the flow rate (Q) was not the only factor to consider in cost estimation of these five plants, a preliminary scatter chart of cost versus inlet flow rate was plotted (Figure 5). The scatter chart did not indicate any obvious function, but was used as a base for further investigation.

The data on these plants was reduced to those factors which, at first glance, would seem to be significant parameters in any cost estimate. Table VII shows these values for the five plants studied.

Note that the pressures are all within the range of 813-1100 psig. and that the temperatures are all within 50°F to 83°F (510° to 543° Rankine). These were the first variables to be ignored (or treated as constants) in this study.

It was recognized at the outset that there could be significant cost differences because of temperature or pressure factors, even within the above range. However, for the first attempt at relating the capital costs of the plants to their operating parameters, these factors were

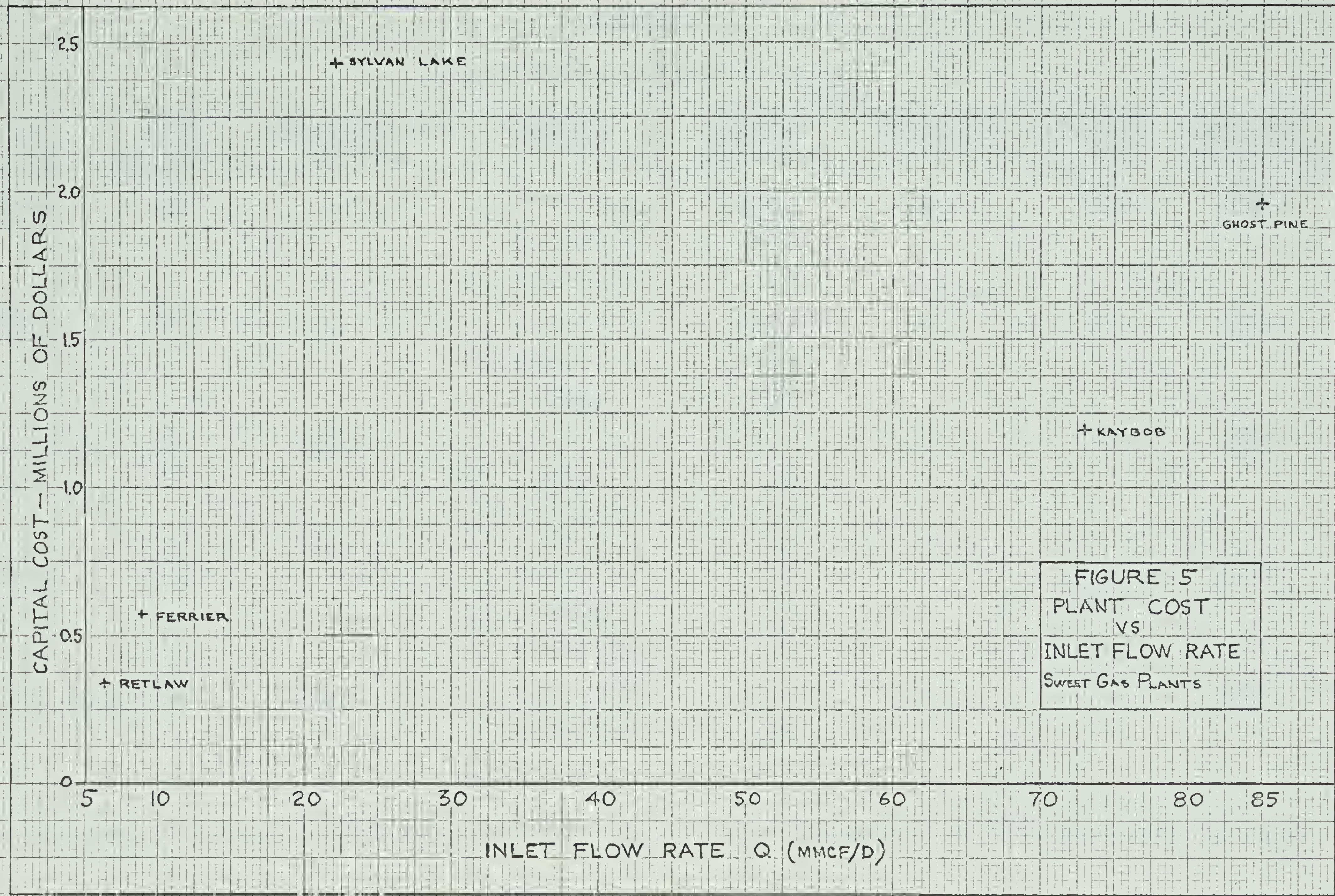

FIGURE 5
PLANT COST
VS
INLET FLOW RATE
SWEET GAS PLANTS

TABLE VII

COST AND PERTINENT OPERATING FACTORS

Plant	Cost Thousands	Inlet Flow Rate MMcf/Day	BBL/MMcf In/Out	Recovery Percentage	Temperature °F	Pressure psig.	Acid Gas Percentage
Ferrier	571	8.9	C_{5+} 58.21/58.13	100.0	60	1100	1.00%
			C_4 21.38/8.30	38.8			
			C_3 43.50/2.57	5.9			
Kaybob	1193	73.0	C_{5+} 4.57/1.99	43.3	50	930	.47%
			C_4 7.48/.19	2.5			
			C_3 17.02/0	nil			
Ghost Pine	1965	85.0	C_{5+} 9.81/7.10	72.4	65	950	.06%
			C_4 11.86/.05	.4			
			C_3 21.70/0	nil			
Retlaw	334	6.95	C_{5+} 5.88/5.03	85.5	83	813	4.20%
			C_4 5.30/.35	7.0			
			C_3 10.05/.09	1.0			
Sylvan Lake	2428	22.1	C_{5+} 20.65/20.46	99.0	60	950	2.96%
			C_4 12.80/12.0	94.0			
			C_3 26.5/23.5	88.7			

held constant as a means of simplifying the analysis. This step was supported by the results of the analysis.

It was assumed that the major variables affecting the cost of a sweet gas plant were: (1) inlet gas flow rate (Q), (2) some function of the liquid content and recovery, and (3) inlet temperature and pressure. Ignoring the last for the reasons cited above, the writer was left with the following general function:

$C = f\ (Q, L, R)$ where,

C = total capital cost of the plant
Q = inlet flow rate, MMcf/D
L = liquid content in gas stream, BBL/MMcf
R = a function of the recovery of these liquids.

During discussions with the Chief Engineer of Pacific Petroleums regarding the costs involved in gas processing plants, the point was raised that possibly a significant variable could be the percentage of liquids recovered from the gas, regardless of the amount of liquids in the stream. Although a cursory examination of the theory involved did not either confirm or deny this thesis, expediency demanded that the first correlations be attempted along this line. Therefore, the general function was reduced to:

$C = f\ (Q, R).$

It was assumed that the heavier liquids are less costly to recover than the lighter liquids. The reasons for this are well founded in theory, as is explained below.

First, as the heavier hydrocarbons will condense before the lighter hydrocarbons, less refrigeration is required to recover them. Thus, a small refrigeration unit, capable of lowering the temperature by only a few degrees, could recover a large percentage of the condensate existing as vapor in the inlet gas, while virtually no propane would be recovered.

Second, if condensate is the only product to be recovered, the equipment required for the transportation and storage of this liquid is relatively simple, but if all three basic natural gas liquids are being recovered, such is not the case. Further processing such as fractionation is required to separate these liquids and the extra equipment required is quite costly.

After investigating the possible use of several functions to plot a cost curve, a simple additive function was decided upon. This function took the form:

$$\text{Capital Cost (as defined in Chapter I)} = F\left(Q\left[AR_{C_{5+}} + BR_{C_4} + CR_{C_3}\right]\right),$$

where A, B, and C are constants and R is the percentage recovery of the natural gas liquids present in the inlet gas stream. A trial and error process was performed by plotting the actual total costs of the five plants under consideration against a function:

$$M_1 = Q\,(AR_{C_{5+}} + BR_{C_4} + CR_{C_3}).$$

The constant A was assumed to be unity, and the values of B and C were varied until the points described by "M_1" and "Capital Cost" fell on a smooth continuous curve. The values of the constants derived in this manner were:

A = 1, B = 1.2, and C = 1.6.

The curve on which the five plants are plotted is shown on Figure 6. One refinement was added at this stage. For plants requiring an amine section[3] the quantity of gas entering the liquids recovery section is not Q, but rather is Qx (100% - acid gas percentage removed). This does not affect any sweet gas plants and so the curve shown on Figure 6 remains the same, but for the general case, the required parameter M_1 is:

$$Q\ (1.0 - a)(1R_{C_{5+}} + 1.2R_{C_4} + 1.6R_{C_3}) \times 10^{-2}$$

where "a" is the decimal fraction of acid gas in the inlet gas stream (provided it is removed by the amine section).

It was recognized that there were possibly other functions which could be used to obtain a logical plot of capital costs versus operating parameters. However, the function derived here was felt to be quite satisfactory and further investigation into other functions was not considered.

Some economic interpretation and support for the curve shown on Figure 6 was warranted. The first term examined was the inlet flow

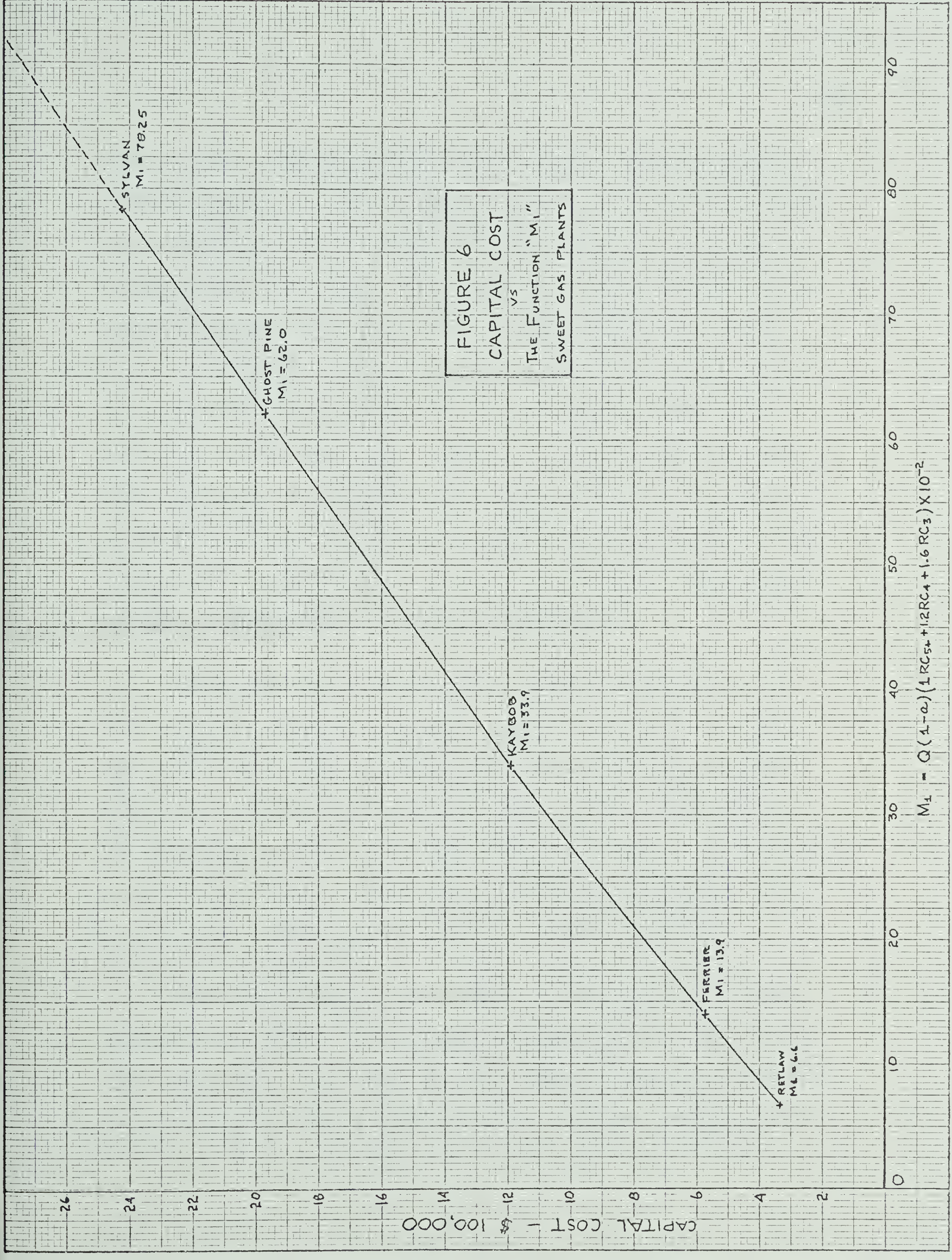

FIGURE 6
CAPITAL COST
vs
THE FUNCTION "M_1"
SWEET GAS PLANTS

rate, or "Q". It is quite obvious that a plant required to process 50 MMcf/DAY would cost more to construct than one required to process only 5 MMcf/DAY. (It is assumed here that all other factors for the two plants are identical.) This assumption is easily verified on the basis of equipment size alone, if the technology involved in both plants is similar.[4] In general terms, a plot of capital cost versus "Q" would show an increase in cost with an increase in "Q" if all other factors were identical. Although the cost per MMcf processed might decline as "Q" increased, the total cost would still increase as the flow rate was increased. It is quite possible that the curve might follow a step function, however in the absence of data suggesting this, a smooth curve is the most logical choice.

The second term to be considered was the recovery factor of the condensate, butanes or propanes. Again (other factors remaining constant) it is logical to assume that a high percentage recovery of butanes, for example, would require greater capital investment than a low percentage recovery as the refrigeration required in the first case is obviously greater. A smooth curve showing increasing cost with increasing percentage recovery provides a rational solution.

The third factor to be considered was most important to the results of this study. It was explained previously that the heavier hydrocarbons were less costly to recover than the lighter hydrocarbons. The problem was to find a weighting factor that would recognize this

difference in cost. The constants (A, B and C) applied to the recovery factors for condensate, butanes and propane accomplish this in that any given recovery percentage is more heavily weighted for the lighter components. With the positive sloping curve suggested above, this weighting automatically increases the capital costs more for the same per cent recovery of the lighter hydrocarbons. Note that it is the ratio between these constants and not their absolute values that is important.

In summary, the curve shown on Figure 6 is logically defensible. It was expected that the curve would flatten at the low end because of minimum size design problems, however recent technological advances in the packaging of these small plants could offset this. Chapter V discusses this aspect more fully.

With the apparent success of this curve plotting, the writer felt that the questions raised by using only the two parameters of inlet gas flow and percentage recovery of liquids should be answered before proceeding with the more complicated plant types. What follows is the result of numerous conversations with several company engineers and is by no means conclusive. However, it was generally accepted by all concerned that while substantiation would be extremely difficult, the statements which follow are entirely feasible.

The problem can be illustrated most clearly by the use of an example. Consider two plants operating under identical conditions

except that in one case the inlet gas contains forty barrels of condensate per million cubic feet of gas and, in the other case, there are only ten barrels of condensate per million cubic feet of inlet gas. (It is assumed that the propane and butanes content of the inlet gas is identical.) According to the function discussed earlier, if the plant process were to recover twenty BBL/MMcf in the first case, and five BBL/MMcf in the second case, the capital costs of the liquid recovery sections of each plant would be identical. To phrase this in general terms, it does not matter what the liquid content of the inlet gas is, rather it is the fraction of this liquid content that is recovered which is the significant cost factor.

A cursory investigation of this statement would suggest that it is in conflict with Postulate 2. The plant with the lower condensate content would obviously require a lower operating temperature than the plant with the higher condensate content. However, the significant point to remember is that lower temperature does not necessarily require a greater refrigeration capacity. This is so because of the two types of heat energy removed in the process. The first type is "sensible heat" which is that heat removed in order to change the temperature of the gas. The second type is "latent heat" which is the heat required to change the phase of the condensate from the gaseous to the liquid phase.

Returning to the example, it is obvious that the <u>sensible</u> heat

required to recover fifty per cent of ten BBL/MMcf is greater than that sensible heat required to recover fifty per cent of forty BBL/MMcf. However, the latent heat required is much greater for the gas having the high pentanes plus content. Thus the two effects tend to cancel each other out.

Theoretically, it would be possible to confirm that this assumption was reasonably valid for the ranges studied, but this would not provide any significant confirmation of the cost function proposed. This is because the piping, processing, and storage requirements would increase as the quantity of liquids removed increased. As the above explanation is plausible, it was not thought necessary to spend the considerable time required to further investigate the theoretical aspects of the assumptions.

Sour Gas Plants

Sulphur plant and stack costs. As mentioned previously, there were very few plants available to the writer with any cost breakdown between the various processing sections. Therefore, no direct cost estimating was possible for the amine section. However, it was felt that these costs could be imputed if all the other processing sections could be costed accurately. As good cost data was available on sulphur plants, this was the next section to be investigated by the writer.

Capital cost data on sulphur recovery plants was provided for

the most part by Black, Sivalls & Bryson, Limited, with the supporting data being provided by Pacific Petroleums' files. The capital cost data provided by B.S. & B. pertains exclusively to their own packaged sulphur plants up to a daily production capacity of 25 long tons per day (2,240 pounds per long ton). Above this figure, their data includes both their own and competitors' projects. The curves derived can be used to estimate the capital costs of both the "In Line" and "Bypass" processes, according to their originator.[5] This view was confirmed in private communication with J.W. Estep of Texas Gulf Sulphur Company. (All plants used in this study are of the "Bypass" type with the exception of the Lone Pine plant, which utilizes the "In Line" process).

Sulphur plant capital costs can be related directly to sulphur production capacity. Note however, that the sulphur plant inlet gas stream contains not only hydrogen sulphide but carbon dioxide and water vapor as well. Therefore the total acid gas inlet flow rate (or the ratio of H_2S to total acid gas) could be used as a parameter in capital cost estimation procedures. In actual practice, however, this factor is ignored, and capital costs are plotted as a function of sulphur production capacity only.[6]

Figure 7 shows two curves of capital cost versus long tons of sulphur produced. The first curve pertains to sulphur plants built prior to 1968, and the second curve is to be used for estimating

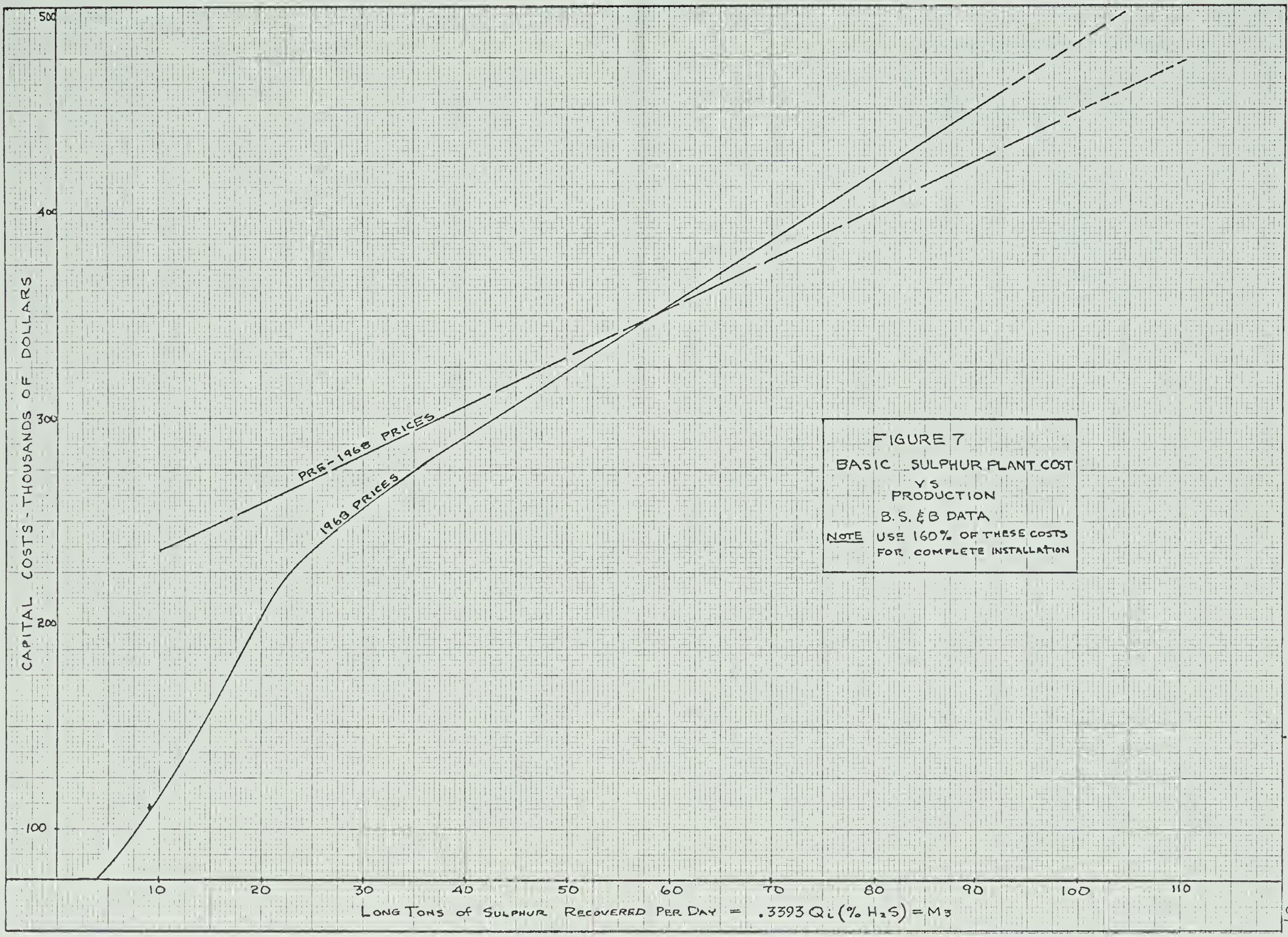

FIGURE 7
BASIC SULPHUR PLANT COST VS PRODUCTION
B.S. & B DATA
NOTE USE 160% OF THESE COSTS FOR COMPLETE INSTALLATION
PRE-1968 PRICES
1968 PRICES
CAPITAL COSTS - THOUSANDS OF DOLLARS
100
200
300
400
500
10
20
30
40
50
60
70
80
90
100
110
LONG TONS of SULPHUR RECOVERED PER DAY = .3393 Qi (% H2S) = M3

future sulphur plant capital costs. Note that the latter curve shows lower capital costs in the lower capacity ranges. This is the result of improved technology and "packaging" techniques in the smaller plants as well as increased price competition in this range. The "tons of sulphur produced per day" has been converted for convenience to a term based on the inlet flow rate (Qi) and hydrogen sulphide content of the gas. There are .377 long tons of sulphur per million cubic feet per one per cent hydrogen sulphide. The sulphur plant efficiency can be assumed at ninety-two per cent. (The normal range is eighty-eight to ninety-six per cent.) There is a further loss in the amine section of about two per cent. Therefore, the overall production of sulphur can be written:

$$P = .9 \times .377 \times Qi \times \% H_2S = .3393\ Qi\ (\% H_2S) = M_3.$$

The cost provided on Figure 7 is the basic cost of the sulphur plant. To this cost must be added a cost estimate for the following items associated with the plant:

Incinerator and Stack, Storage Facilities,
Startup Costs, Operators' Overhead,
Site Preparation and Foundations,
Sales Tax on Equipment, and
Auxiliary Piping and Electrical Facilities.

The stack costs represent a large portion of this additional cost and, unfortunately this cannot be linked directly to the plant

operation. The Alberta Government Department of Health has close control over stack height and the variables they consider are plant operating parameters, locale, wind patterns and other meteorological factors, nearby land elevations, the presence of population centres, and local agricultural activities.

Therefore it is almost impossible to provide a good estimate of stack height based on operating parameters alone. However, most sulphur plants have a stack height of one hundred to two hundred feet, and as Figure 8 indicates, the percentage error in the total plant costs would not be serious even if the height estimation were greatly in error.

Stankiewicz derived capital cost curves for stacks based on the amount of steel required for various heights and diameters.[7] Individual curves for the stack, supports, foundations and lining were illustrated. However, this method of estimation still requires stack height and diameter data which, as noted previously, is not solely dependent on operating parameters.

One important point should be noted regarding stack costs. When a sulphur plant is not installed, a stack is still required to incinerate any hydrogen sulphide produced from the amine section.

The extra costs noted previously, added to the basic costs as quoted by Black, Sivalls & Bryson, totaled approximately one

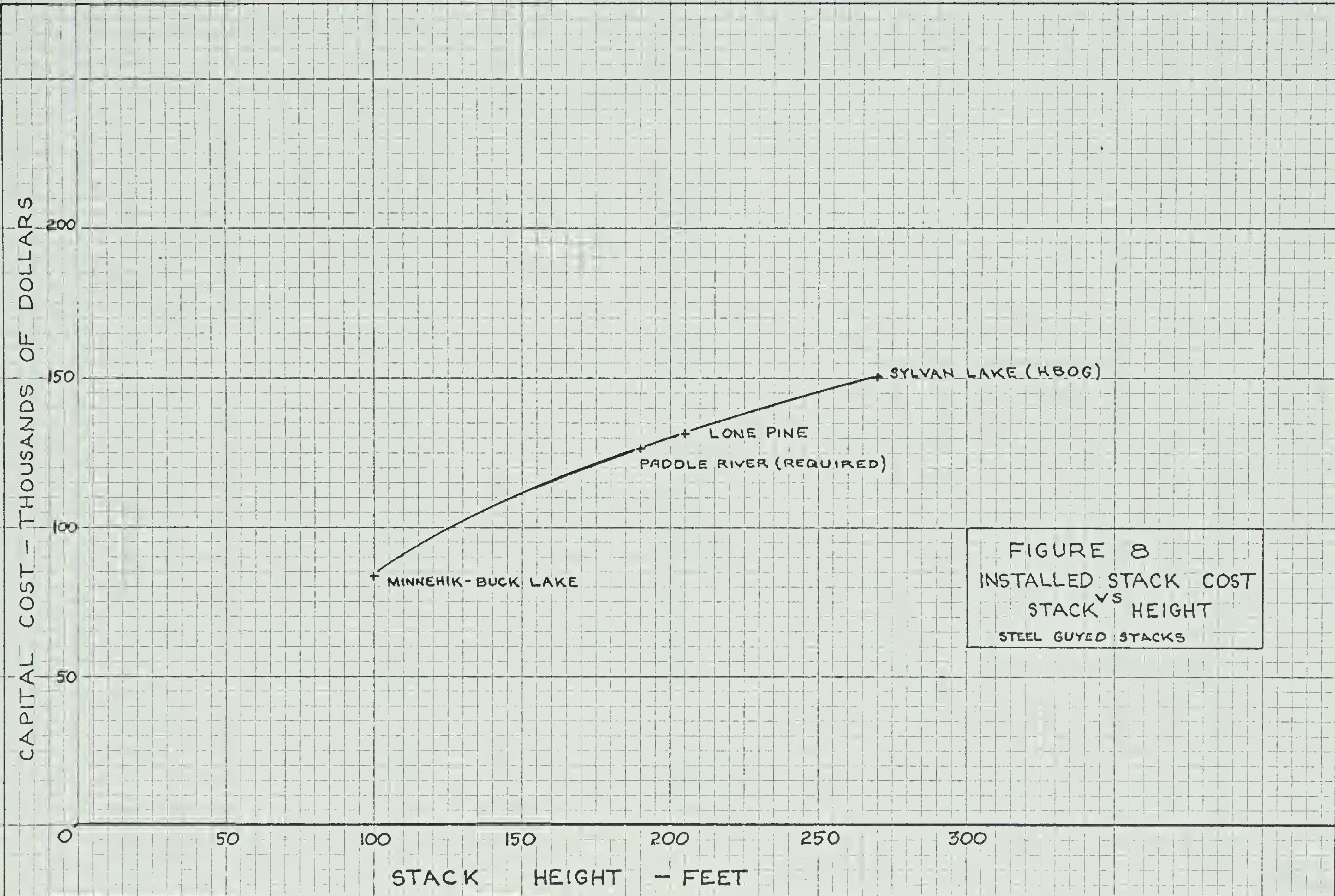
FIGURE 8
INSTALLED STACK COST
STACK VS HEIGHT
STEEL GUYED STACKS
CAPITAL COST - THOUSANDS OF DOLLARS
STACK HEIGHT - FEET
0
50
100
150
200
250
300
SYLVAN LAKE (H.B.O.G.)
LONE PINE
PADDLE RIVER (REQUIRED)
MINNEHIK-BUCK LAKE

hundred and sixty per cent of the quoted costs on three sulphur plants on which information was available to the writer. Therefore, this factor was used as a multiplier for the purposes of estimating the total installed sulphur plant costs in this study.

Shown below is the comparison between the cost of the three sulphur recovery plants as estimated using Figure 7, and the actual costs of these three plants. Note that only the Minnehik-Buck Lake plant was built by B.S. & B. This tends to verify the adequacy of their curves.

Plant	Theoretical Production LT/Day	Design Production* LT/Day	Cost Shown In Figure 7 Thousands	x 1.6 Thousands	Actual Completed Cost Thousands
Minnehik-Buck Lake	20.50	26.0	273.0	437	445
Sylvan Lake**	13.7	10.0	237.0	379	379
Lone Pine	99.5	112.0	474.0	759	740

*The design production rate can be higher or lower than the theoretical production rate, depending on expected future inlet flow rates and forecasted hydrogen sulphide production. These plants are often slightly overdesigned.

**Sylvan Lake Plant #2, operated by Hudson's Bay Oil & Gas Ltd.

Amine section costs. As was mentioned earlier, the only available method of extracting the amine section costs from the total plant costs was to assume that these costs were residual costs after all other costs had been tabulated. The method used was as follows.

First, the liquid section recovery cost chart (Figure 6) was assumed to be valid for plants processing both sweet and sour gas. Then the sulphur plant costs and stack costs were added where applicable, and the results were totaled. The difference between these total chart costs and the actual adjusted total costs of the plants was assumed to be attributable to the costs of the amine section. Table VIII shows the tabulation and the resulting imputed amine section cost for the five sour gas plants investigated.

It should be noted that this method of arriving at amine section costs automatically transferred all of the possible error in the estimation of other section costs to this amine estimate. An error quite possibly was made in the estimation of the liquid recovery section, the sulpur plant or the stack, but because of the assumptions made, no error was evident in these sections. Therefore, it was to be expected that these combined errors could result in a very poor correlation on the amine section cost curves.

There were two obvious parameters to consider in the cost estimation of this section. One, of course, was the inlet flow rate, and the other was the percentage of acid gas contained in the inlet stream. Multiplying these parameters together resulted in the actual flow rate (in MMcf/Day) of acid gas to be removed. This seemed to be a logical choice to the writer, and therefore the imputed amine section costs were plotted against the function M_2, where M_2 equalled

TABLE VIII

DERIVATION OF AMINE SECTION COSTS
(All Costs in Thousands)

Plant	Liquid Section M_1	Liquid Section Fig. 6 Cost	Sulphur Section 160% of Fig. 7 Costs	Stack Costs Figure 8	Total Chart Cost	Actual Adjusted Plant Costs	Actual Costs Less Chart Cost = Amine Section Cost
Minnehik-Buck Lake	61.6	1960	437	nil	2397	3143	746
Paddle River	27.8	1010	nil	126	1136	1716	580
Gilby	35.9	1240	nil	nil	1240	1617	377
Lone Pine	30.9	1100	759	nil	1859	3160	1301
Boundary Lake	52.8	1712	nil	nil	1712	2089	377

the inlet flow rate times the percentage of acid gas in the stream. This data was plotted on Figure 9.

The plot shown on Figure 9 would suggest that a plant processing 150 MMcf/D of acid gas would cost approximately the same to build as a plant processing only 0.5 MMcf/D, while a plant processing 67.1 MMcf/D would cost considerably more. This is not logically defensible if one assumes that the acid gas processed is the only major variable affecting size, and hence cost of an amine unit. (The validity of this assumption is discussed later in this thesis.)

Due to the rather unsatisfactory nature of the plot on Figure 9, the operators of all five sour gas plants were contacted and the files re-examined in an attempt to find a possible reason for the variance.

First, the cost differences between the MEA and DEA processes were investigated. (The Boundary Lake and Minnehik-Buck Lake plants use MEA while the Paddle River and Lone Pine plants use DEA.) Discussions with process engineers revealed that in some cases, the DEA process could be slightly cheaper, but that the <u>maximum</u> capital cost difference between the two processes would be about 10%.[8] Due to the small estimated difference and the uncertainty of the estimate, it was decided to ignore the differences in capital costs of the two processes for the purposes of this paper.

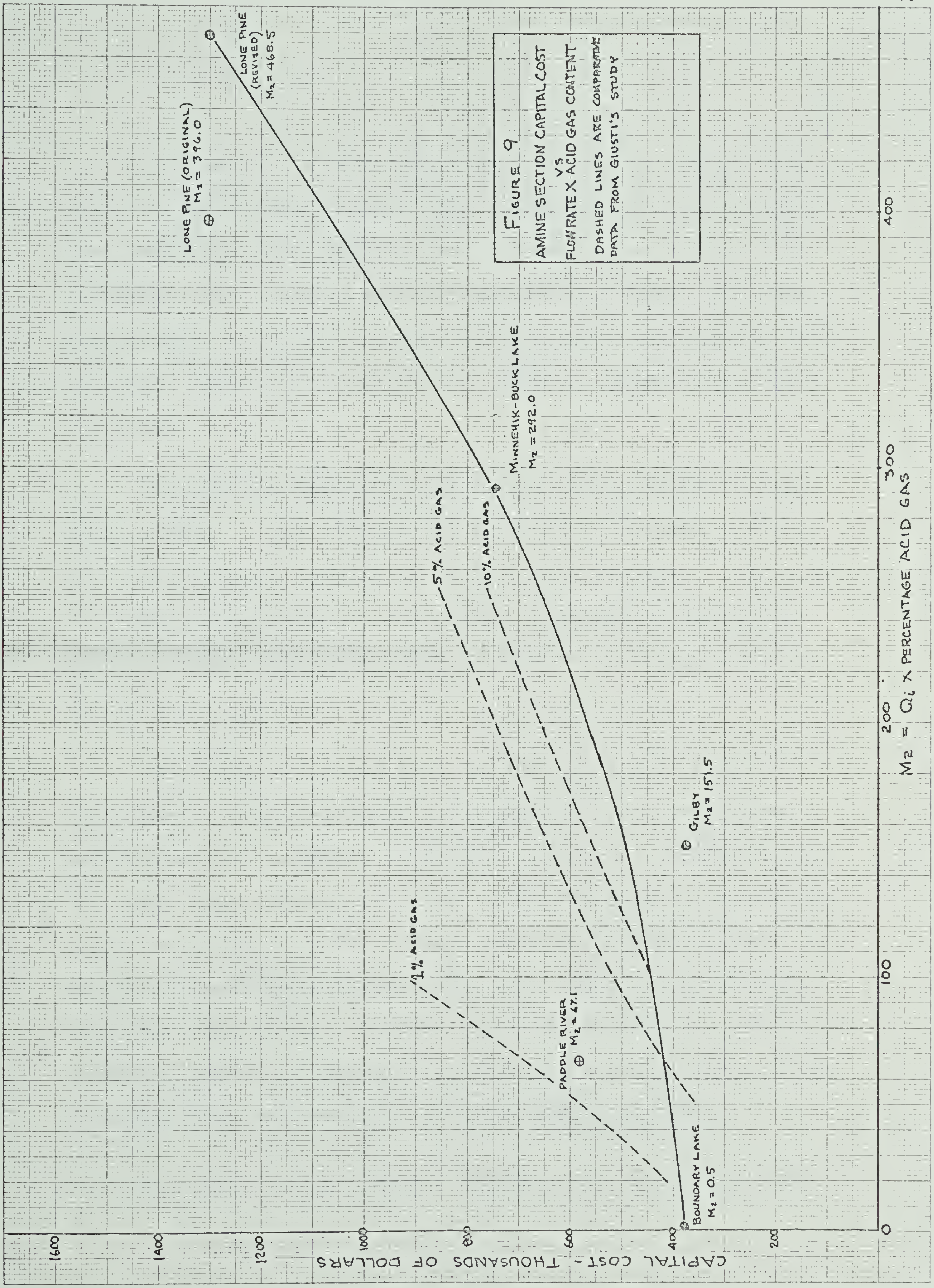

FIGURE 9

AMINE SECTION CAPITAL COST VS FLOW RATE X ACID GAS CONTENT

DASHED LINES ARE COMPARATIVE DATA FROM GIUSTI'S STUDY

The Texaco Gilby plant was the only dry dessicant plant investigated for the purposes of this study. A normal characteristic of this type of plant is lower capital costs and higher operating costs. Although this was recognized at the beginning of the study, this plant was included, with the reservation that the capital cost would probably be lower. This was the case, and this plant was ignored when fitting the curve.

The Lone Pine plant was found to have excess capacity in the amine section. It was revealed that the plant was designed to handle up to 12.5% hydrogen sulphide at the same inlet flow rate. (Note also the excess sulphur capacity of this plant.) Therefore, the function M_2 was increased from 396.0 to 468.5. This had the effect of flattening the curve somewhat.

There were two reasons noted for the high cost of the Paddle River plant. First, this plant was designed for unattended operation and thus had more automatic instrumentation and control equipment than the normal plant. Second, the stack constructed at this plant was not a steel guyed stack; it was of brick construction. This increases the stack cost considerably however, the overall plant cost shows only a small percentage change. The first reason would change the cost considerably however, and therefore the cost of a normally operated plant in Paddle River was considered to be lower than that shown.

The curve shown on Figure 9 thus has a weak statistical base.

The unfortunate lack of data precluded any further verification. However, the curve was considered to verify the expected trend and to show at least an approximately correct magnitude. The Boundary Lake plant cost data in the files gave an excellent breakdown, as mentioned previously, and this cost verified the imputed cost in this study, thus providing one excellent point on the chart. Minnehik-Buck Lake costs were discussed with the operator and it was verified verbally that the amine plant costs were approximately three quarters of a million dollars. No adequate cost breakdown was available for the Lone Pine plant.

For comparison, Figure 9 shows the results of the study by Gino Giusti referred to in Chapter III.[9] Giusti assumed that both the inlet flow rate and the acid gas content were independent variables and constructed a series of curves of capital cost versus flow rate for various gas concentrations. These curves, adjusted for price level change, appear as the dashed lines on Figure 9.

Estep made basically the same assumption.[10] He derived a capital cost curve using a fixed flow rate and plotting capital cost versus acid gas content. Unfortunately, no curves for different flow rates were presented.

Giusti's assumption was initially thought to be valid and it was unfortunate that neither Giusti's article or the data available for this study could be used to support his argument. This writer recog-

nized that the amine section costs could be influenced independently by both the inlet flow rate and the acid gas content but <u>only</u> in the contactor. (See Figure 2.) Once the rich amine leaves the contactor, these two parameters cease to act independently and their product (or the flow rate of acid gas) is the valid measure of size. As most of the cost involved in this section is related to the regenerator and its associated equipment, it was felt that the error induced by using the parameter M_2 was negligible, except perhaps at the extreme size limits of the process. (This was not supported by the Boundary Lake data.)

Note however that the maximum acid gas content of the plants studied was under 20%. It is entirely possible that Giusti's assumptions should be considered when estimating the capital costs of plants processing a much higher percentage of acid gas in the inlet stream.

This completes the study on the capital costs of natural gas processing plants. A series of curves have been derived by which the gas processing engineer or executive can rapidly estimate the costs of each pertinent section of any conventional gas processing plant. The final chapter of this thesis tabulates the results of the study and shows the error obtaining when the suggested estimation method is used. As a further check on the validity of the method, the actual costs and detailed estimates of several other plants are compared with the estimates arrived at by the use of this method.

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[1]This index is published weekly in Engineering News Record.

[2]Published the first week of each month in the Oil and Gas Journal.

[3]Generally it is quite easy to decide if the plant will require an amine section. For example, the presence of all but trace amounts of hydrogen sulphide will, in most cases, indicate that a sweetening process will be required.

[4]Except for the improvements in packaged design of very small plants, the technology involved in liquids recovery has remained virtually unchanged in the past 10 years.

[5]Private communication with Mr. D.R. Henderson, Black, Sivalls and Bryson Limited, Calgary, Alberta.

[6]James W. Estep, op. cit., p. 27.

[7]E.J. Stankiewicz, "How to Estimate Stack Costs", Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 253-258.

[8]Private communication with Mr. James W. Estep of Texas Gulf Sulphur Company, Calgary; Mr. K.J. Fitzgerald of Barry and Richardson, Calgary.

[9]Gino P. Giusti, "Sulphur Recovery Processes", Oil and Gas Journal, February 22, 1965, p.99

[10]James W. Estep, op. cit., pp. 3, 25, 26.

CHAPTER V

SUMMARY AND CONCLUSIONS

I. INTRODUCTION

The purpose of this chapter is twofold. First, the parameters used in estimating the ten gas plants are tabulated and the curves derived in the previous chapters are used to estimate the total capital costs of the plants studied. It is necessary to stress that the points plotted on these curves represent estimates and do not necessarily reflect the actual cost of the plants.

Second, several additional plant cost estimates are provided for validation purposes. The raw data on these plants is presented in Appendix C, and their estimates are also plotted on the curves referred to above.

The results of the cost estimation procedure both for the original plants and the additional plants are compared with the respective adjusted actual costs of the plants involved. Percentage errors are computed and discussed. The chapter concludes with a discussion of the overall results of the study, and suggestions for further work in this field.

II. SUMMARY OF INITIAL PLANT ESTIMATES

Table IX shows the functions M_1, M_2 and M_3 for the ten gas plants initially studied, and summarizes the resulting estimates. Figures 10,

11, and 12 provide cost curves for the liquids recovery section, the amine section, and the sulphur recovery section respectively. As there is only one plant within the scope of this study that requires a stack cost estimate, this chart is not reproduced here. The reader is referred to Figure 8 for verification of this cost.

III. ADDITIONAL PLANTS ESTIMATED BY THIS METHOD

Appendix C shows the required data for several additional estimates used to verify the accuracy of the proposed method. Generally speaking, the data available on these plants was not as good as that collected for the original ten plants. Some of the problems involved are discussed below.

Hudson's Bay Oil and Gas Ltd. - Sylvan Lake Plant. The existing Sylvan Lake plant has deep cut facilities installed and the parameter M_1 is too large to fit within the bounds of this study. It was decided to estimate the original (shallow cut) facility and the data in Appendix C deals only with this original plant. The obsolete brick stack was assumed to be replaced by a comparable steel stack for estimation purposes. This was probably the best plant data available of the six additional plants studied.

Clive Alix Plant - Hydrogas Ltd. This plant has yet to be constructed but the cost data provided is in the form of a detailed design estimate submitted by the consulting firm of Barry and Richard-

TABLE IX

SUMMARY OF PARAMETERS AND ESTIMATES
(All Costs in Thousands)

Plant	Liquid Section Cost		Amine Section Cost		Sulphur Plant Cost		Stack Cost	Total Estimated Cost
	M_1	Estimate	M_2	Estimate	M_3	160% x Chart Cost		
Ferrier	13.9	$ 571	nil	$ nil	nil	$ nil	$ nil	$ 571
Kaybob	33.9	1193	nil	nil	nil	nil	nil	1193
Ghost Pine	62.0	1965	nil	nil	nil	nil	nil	1965
Retlaw	6.6	334	nil	nil	nil	nil	nil	334
Sylvan Lake	78.3	2428	nil	nil	nil	nil	nil	2428
Minnehik-Buck Lake	61.6	1960	292.0	746	26	437	S*	3143
Paddle River	27.8	1010	67.1	420	nil	nil	126	1556
Gilby	35.9	1240	151.5	492	nil	nil	nil	1732
Lone Pine	30.9	1100	468.5	1301	112	759	S	3160
Boundary Lake	52.8	1712	.5	377	nil	nil	nil	2089

*"S" denotes stack costs that are included in the sulphur plant costs.

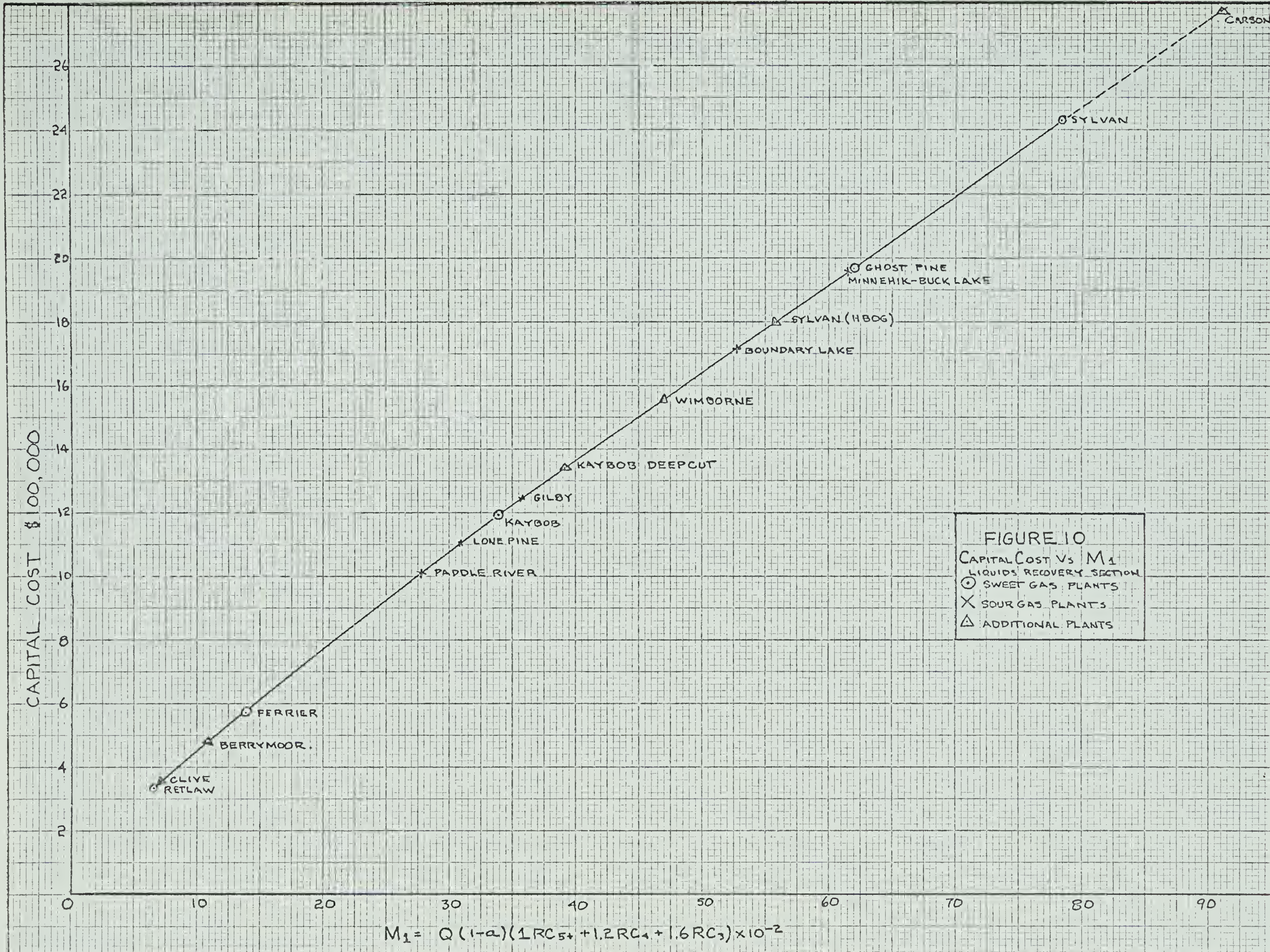

FIGURE 10
CAPITAL COST Vs M_1
LIQUIDS RECOVERY SECTION
SWEET GAS PLANTS
SOUR GAS PLANTS
ADDITIONAL PLANTS
CAPITAL COST $100,000
$M_1 = Q(1-a)(1RC_{5+} + 1.2RC_4 + 1.6RC_3) \times 10^{-2}$
CARSON
SYLVAN
GHOST PINE
MINNEHIK-BUCK LAKE
SYLVAN (HBOG)
BOUNDARY LAKE
WIMBORNE
KAYBOB DEEPCUT
GILBY
KAYBOB
LONE PINE
PADDLE RIVER
FERRIER
BERRYMOOR.
CLIVE
RETLAW
0
10
20
30
40
50
60
70
80
90
2
4
6
8
10
12
14
16
18
20
22
24
26

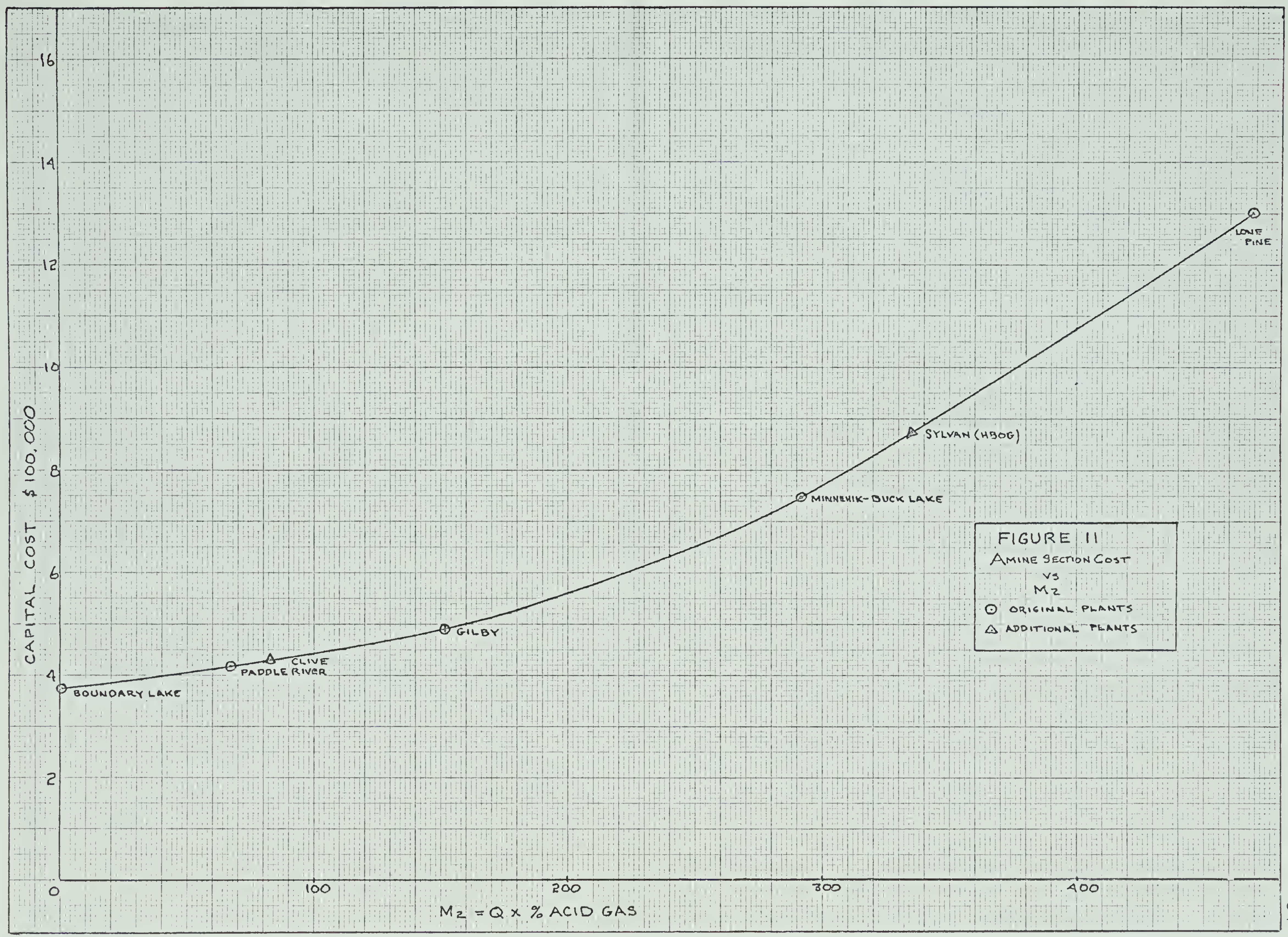
FIGURE 11
AMINE SECTION COST
vs
M2
ORIGINAL PLANTS
ADDITIONAL PLANTS
CAPITAL COST $100,000
M2 = Q x % ACID GAS
BOUNDARY LAKE
PADDLE RIVER
CLIVE
GILBY
MINNEHIK-BUCK LAKE
SYLVAN (H3OG)
LONE PINE
0
2
4
6
8
10
12
14
16
100
200
300
400

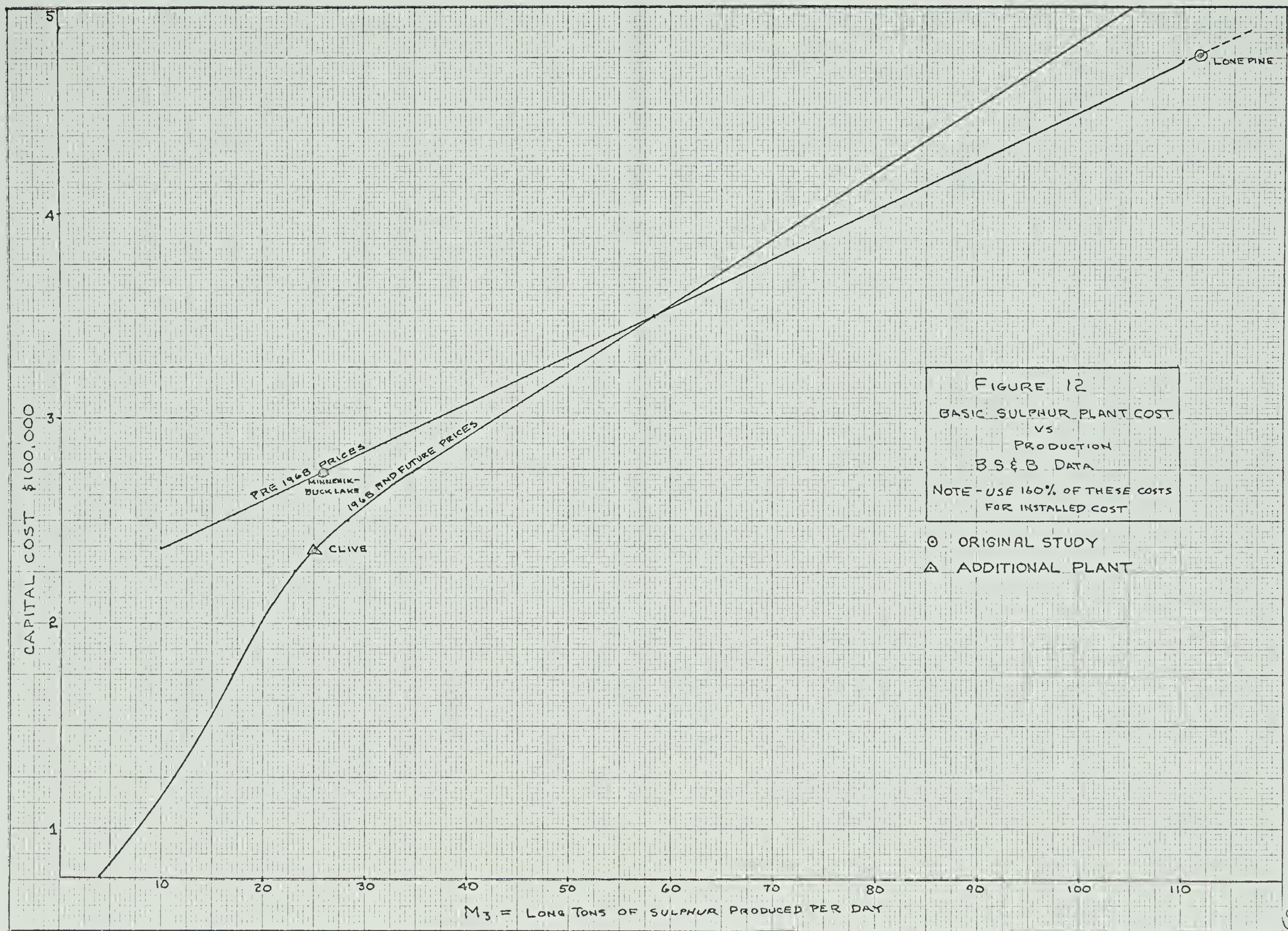
FIGURE 12
BASIC SULPHUR PLANT COST
VS
PRODUCTION
B S & B DATA
NOTE - USE 160% OF THESE COSTS
FOR INSTALLED COST
ORIGINAL STUDY
ADDITIONAL PLANT
LONE PINE
PRE 1968 PRICES
MINNEHIK-
BUCK LAKE
1968 AND FUTURE PRICES
CLIVE
CAPITAL COST $100,000
1
2
3
4
5
10
20
30
40
50
60
70
80
90
100
110
M3 = LONG TONS OF SULPHUR PRODUCED PER DAY

son. As these detailed cost estimates are usually within plus or minus five per cent, the data was felt to be highly satisfactory for the writer's purposes.

Kaybob Deep Cut Facilities - Pacific Petroleums Ltd. This data also pertains to an unconstructed gas plant. It was provided in the form of a proposal by K.C.M. Engineering Ltd. to the writer's employer. Although a detailed design was not done, the data provided by the proposal was felt to be realistic by the company engineers and therefore it was included in this study.

Carson Creek Shallow Cut Facility - Mobil Oil Canada Ltd. This plant was the one eliminated from the original study because of its high operating pressures. It was not expected that this plant could be estimated by the proposed method, but nevertheless, it was included for comparison purposes.

Wimborne Plant - Mobil Oil Canada Ltd. This was probably the most unsatisfactory plant data collected. Discussions with Mobil personnel revealed only two details on cost; first, that the total cost was approximately five million dollars, and second, that the percentage of these costs attributable to the sulphur recovery and amine sections was about seventy per cent. It was recognized that this type of data would not substantially verify the proposed estimation method, however it was included for comparison purposes.

Berrymoor (Pembina) Plant - Pan American Petroleum Corporation. This is a small simple refrigeration type plant. One possible source of error in the data provided was in the condensate recovery percentage. As this plant has not been operating at capacity for several years, it was felt that the figure given might not be representative of the recovery under full load. Generally speaking however, the data received was thought to be satisfactory for the purposes of this paper.

The cost estimation parameters M_1, M_2 and M_3 for these six plants are calculated in Appendix C and the resulting cost estimates are plotted on Figures 10, 11 and 12.

IV. EXAMINATION AND DISCUSSION OF THE ESTIMATES

Table X summarizes the results of the proposed estimation procedure for the 16 plants studied and shows the percentage error resulting from the estimate. (The percentage error was computed using the actual cost as a base.)

Eight of the original ten plants were used to derive the cost curves and therefore the estimates show no error, as expected. The error inherent in both the Gilby and Paddle River plants was discussed in Chapter IV and no further comment is necessary here. Note that both estimates are well within the twenty per cent accuracy requirement of this study.

TABLE X

COMPARISON OF ESTIMATES TO ACTUAL ADJUSTED COSTS
(All Costs in Thousands)

Plant	Section Capital Cost Estimates				Total Estimate	Actual Costs	Percentage Error
	Liquids	Amine	Sulphur	Stack			
Retlaw	334	nil	nil	nil	334	334	nil
Ferrier	571	nil	nil	nil	571	571	nil
Kaybob	1193	nil	nil	nil	1193	1193	nil
Ghost Pine	1965	nil	nil	nil	1965	1965	nil
Sylvan Lake	2428	nil	nil	nil	2428	2428	nil
Minnehik-Buck Lake	1960	746	437	nil	3143	3143	nil
Paddle River	1010	420	nil	126	1556	1716	minus 9%
Gilby	1240	492	nil	nil	1732	1617	plus 7%
Lone Pine	1100	1301	759	nil	3160	3160	nil
Boundary Lake	1712	377	nil	nil	2089	2089	nil
Sylvan - H.B.O.G.	1800	875	nil	nil	2675	2950	minus 9%
Clive	355	430	376	nil	1161	1319	minus 12%
Kaybob Deep Cut	1340	nil	nil	nil	1340	1290	plus 4%
Carson Creek	2770	nil	nil	nil	2770	2698	plus 3%
Wimborne	1555	nil	nil	nil	1555	1844	minus 16%
Berrymoor	480	nil	nil	nil	480	351	plus 37%

The Sylvan Lake (H.B.O.G.) plant was also within limits despite the lack of a satisfactory plant cost breakdown. It was felt that, within the total cost provided, there could be substantial costs that pertained to equipment outside the scope of this study. (This was observed in several other cases.)

Both the Clive plant and the Kaybob deep cut plant were examples of more sophisticated estimating procedures and the apparent error in this writer's method is not important enough to warrant discussion.

The surprising accuracy of the Carson Creek estimate prompted the writer to discuss this plant in detail with a process engineer. It was discovered that only the inlet separators were under the high pressure initially attributed to the whole plant, and that the rest of the plant operated at normal pressures. Therefore this plant definitely fit within the scope of the study and a reasonably close estimate was to be expected.

The Wimborne plant estimate was actually compared to a rough guess only, and little validation of the writer's method was gained by its use. However, the two estimates were within the study limits.

The Pan American Berrymoor plant presented the only serious question as to the success of this study. The estimate was, according to the data supplied, $130,000 or 37% in excess of the actual plant cost.

The writer believes that this error could have been caused by any one of the following factors: (1) error in the data supplied, (2) contractor error, or (3) a breakdown of the cost relationship postulated, at the low end of the curve.

First, this was only a verbal quote. Although the inlet flow (Q) was verified elsewhere,[1] the recovery factor and the cost could quite easily have been in error. The cost estimate itself could have been inaccurate in that it did not reflect the total expenditures required to put the plant on stream. Unfortunately, no further data was available to the writer.

Second, this cost could have been the result of contractor error, in that the plant should have cost considerably more.

The writer concedes that the cost relationships assumed in Figure 10 may not be accurate in the low ranges. There could be some overlooked factor in this range that could have a significant effect on capital costs. For example, this factor could be the result of improvements made in small "packaged" plants constructed recently. (Some process engineers feel that this has resulted in <u>reduced</u> capital costs for small simple refrigeration plants constructed in the past two or three years.)

The success of the method in estimating the costs of the Retlaw plant does not support this too strongly, however, as this plant was

built only one year prior to the Berrymoor plant and is very similar in design. Furthermore, the Ferrier (shallow cut) and Clive (deep cut) estimates were both quite satisfactory, which would tend to verify the adequacy of the curves in the low ranges.

The Berrymoor estimate suggests one area of possible weakness in this study. Nevertheless, the writer does not consider that this one unsatisfactory estimate negates the overall success of the study.

The writer believes that this study has accomplished its objective, first, because the cost relationships suggested are theoretically defensible, and second, because the range of error resulting from the estimation method is quite low. The average error for the sixteen plants is only 6.0% while the average error resulting for those plants not used to derive the curves was only 12.1% as compared to the objective of plus or minus 20% accuracy.

V. CONCLUSION

This paper has developed a method of preliminary capital cost estimation for use in estimating small to medium sized natural gas processing plants situated in Alberta. The accuracy of this method has been shown to be generally superior to the accuracy attributed to those preliminary estimation methods currently in use today.

While this method is apparently successful in rapidly estimating the capital costs of gas plants, there is much room for further

work in this field. The examination of very small gas plants and their operating parameters would do much toward improving the accuracy of this method in these ranges. Similarly, the curves could be extended to include larger plants. Further data on plants within the scope of the study would be extremely useful in verifying or modifying the cost curves derived.

Of paramount importance in any further work is the availability of more and better data on all aspects of gas processing plant construction and operation. It is felt that until the owners of the many Alberta gas plants are convinced that the free exchange of data would be for the common good, improvement of this or any other method will be extremely slow.

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[1] "CPE's 5th Census of Canadian Gas Processing Plants", Canadian Petroleum, July 1967, pp. 64-66.

BIBLIOGRAPHY

BIBLIOGRAPHY

A. BOOKS

Aries, Robert S., and Robert D. Newton. Chemical Engineering Cost Estimation. New York: McGraw-Hill Book Company, Inc., 1955.

Katz, Donald L., et al. Handbook of Natural Gas Engineering. New York: McGraw-Hill Book Company, Inc., 1959.

Kirkbride, Chalmer G. Chemical Engineering Fundamentals. New York: McGraw-Hill Book Company, Inc., 1949.

Natural Gas Processors Suppliers Association. Engineering Data Book. Eighth edition. Tulsa, Oklahoma: Natural Gas Processors Suppliers Association, 1966.

Perry, John H., C.H. Chilton, S.D. Kirkpatrick. Chemical Engineers Handbook. Fourth edition. New York: McGraw-Hill Book Company, Inc., 1963.

Peters, Max S. Plant Design and Economics for Chemical Engineers. McGraw-Hill Series in Chemical Engineering. New York: McGraw-Hill Book Company, Inc., 1958.

Petroleum Industry Training Service. Gas Processing. Second edition. Calgary, Alberta: Petroleum Industry Training Service, 1968.

Schweyer, H.E. Process Engineering Economics. New York: McGraw-Hill Book Company, Inc., 1955.

Zimmerman, O.T., and I. Lavine. Chemical Engineering Costs. Dover, New Hampshire: Industrial Research Service, 1950.

B. ARTICLES

Chilton, C.H., "Six-Tenths Factor Applies to Complete Plant Costs," Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 282-284.

Giusti, Gino P., "Sulphur Recovery Processes," Oil and Gas Journal, February 22, 1965, pp. 99-104.

Hilborn, J.D., (ed.). "CPE's 5th Census of Canadian Gas Processing Plants," Canadian Petroleum, July 1967, pp. 64-66.

Hillsman, J.M., "Faster Cost Estimates for Gas Plants," Hydrocarbon Processing and Petroleum Refiner, Vol. 42, No. 4, April 1963, pp. 171-174.

Lang, H.J., "Cost Relationships in Preliminary Cost Estimation," Chemical Engineering, October 1947, p. 117.

Lang, Hans J., "Simplified Approach to Preliminary Cost Estimation," Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 12-14.

Stankiewicz, E.J., "How to Estimate Stack Costs," Cost Engineering in the Process Industries, Cecil Chilton, Editor (New York: McGraw-Hill Book Company, 1960) pp. 253-258.

Weber, G., (ed.). "Nelson Cost Indexes," The Oil and Gas Journal, November 6, 1967, p. 106.

Wessel, H.R., "How to Estimate Costs in a Hurry," Chemical Engineering, January 1953, p. 168.

C. PAPERS

Estep, James W., and Edward W. Plum, "Economics of Sour Gas Industry," (Paper read at the Fall Meeting of the Society of Mining Engineers, Los Vegas, Nevada, October 1967.)

APPENDICES

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Ferrier

Gas Type: wet solution gas, oil slugs, (2 phase flow)

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	8.89	8.0
Pressure PSI	1100	950
Temperature °F	60	110° max.
Water Content	saturated	4 #/mm

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	75.27		82.72	liquid stream	
C_2	10.54		10.54		
C_3	5.99	43.50	4.23	2.57	5.9
C_4	2.55	21.38	.84	8.30	38.8
C_{5+}	4.23	58.21	.12	58.13	100.0
N_2	.42		.46		
CO_2	1.00		1.09		
H_2S	nil		nil		

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)

3 stage flash separation - refrigeration

Actual Total Cost: Built 1964

Cost - $462,000

Cost Breakdown: (If available)

$414,000 plant

$ 48,000 pipeline

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Kaybob Notikewan and Beaverhill Lake

Gas Type: B.H.L. - solution gas; NOT - non associated gas

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	73.0	72.5
Pressure PSI	930	930
Temperature ^{o}F	50^{o}	10^{o}
Water Content	saturated	4 #/mm

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	88.748		89.21	liquid stream	
C_2	5.697		5.65		
C_3	2.577	17.02	2.47		nil
C_4	.975	7.48	.86	.19	2.5
C_{5+}	.484	4.57	.28	1.99	43.4
N_2	1.050		1.06		
CO_2	.469		.47		

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed) shallow cut (refrigeration) process

Actual Total Cost: $920,000

Cost Breakdown: (If available)

original plant	(1962)	- $400,000
expansion	(1966)	- 520,000

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Minnehik-Buck Lake   Gas Type: wet, sour

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	66.38	60.63
Pressure PSI	973	890
Temperature °F	64°	122°
Water Content	saturated	-5° D.P.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	82.04		88.31	liquids stream	
C_2	7.05		7.06		
C_3	2.68	18.92	2.52	.02	nil
C_4	1.37	11.18	1.07	1.94	17.3
C_{5+}	1.83	20.61	.49	15.72	76.3
N_2	.50		.53		
CO_2	3.62		.02		
H_2S	.91				

Sulphur Data:

Sulphur Production - 22.77 LT/D
Recovery Factor - .9
Estimated Actual Production - 20.50 (plant sized at 26 LT/D)

Cost of Plant: (Note any deviations from basic design processes assumed)

MEA acid gas removal, 2 stage sulphur recovery, refrigeration

Actual Total Cost: $ 2,296,000

Cost Breakdown: (If available)

original plant	(1961) -	$1,170,000
expansion	(1965) -	681,000
sulphur plant	(1967) -	445,000

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Boundary Lake   Gas Type: slightly sour, wet solution gas.

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	17.82	15.94
Pressure PSI	55	680, after compression, 867 & 70°F.
Temperature °F	30	97
Water Content	saturated	-20°F

Gas Analysis:	MOL%	BBL/MMcf Residue	MOL%	BBL/MMCF Residue	Recovery %
C_1	82.44		88.32		
C_2	9.19		9.68	liquid stream	
C_3	5.15	37.89	1.91	22.54	59.5
C_4	2.20	18.65	.09	16.49	88.4
C_{5+}	.99	9.97		9.48	95.0
N_2	nil				
CO_2	.03				
H_2S	trace				

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)
gas turbine use

Actual Total Cost: $1,525,000 excluding compression costs

Cost Breakdown: (If available) amine section cost - $380,000

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Paddle River   Gas Type: sour solution gas

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	28.8	26.056
Pressure PSI	774	720
Temperature °F	75	93
Water Content	saturated	-30° D.P.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	84.955		87.73		
C_2	7.388		7.557	liquid stream	
C_3	2.698	18.57	2.748	0	0
C_4	.897	4.48	.840	.535	11.94
C_{5+}	.812	7.99	.139	6.746	84.4
N_2	.920		.950		
CO_2	1.939				
H_2S	.391				

Sulphur Data: Sulphur Production - 4.24   Recovery Factor - .90
Estimated Actual Production - 3.82, no sulphur plant now

Cost of Plant: (Note any deviations from basic design processes assumed)
refrigeration 10°F, 190 foot stack, unattended operation

Actual Total Cost: 1966 - $1,546,000

Cost Breakdown: (If available) none

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Ghost Pine   Gas Type: wet

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	85.0	82
Pressure PSI	950	915
Temperature ^{o}F	65	60
Water Content	saturated	$15^{o}F$ D.P.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	85.42		86.08		
C_2	7.16		7.19	liquid stream	
C_3	3.28	21.70	3.28	0	0
C_4	1.55	11.86	1.54	05	0.4
C_{5+}	1.00	9.81	.30	7.10	72.4
N_2	1.53		1.54		
CO_2	.06		.07		

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)
standard

Actual Total Cost: (1967) $1,965,000

Cost Breakdown: (If available) none

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Retlaw   Gas Type: wet - slightly sour

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	6.945	6.770
Pressure PSI	813	758
Temperature $^{\circ}F$	83	115
Water Content	saturated	$+3^{\circ}F$ D.P.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	82.74		84.20		
C_2	5.35		5.32	liquid stream	
C_3	1.50	10.05	1.42	.09	1%
C_4	.68	5.30	.61	.35	7%
C_{5+}	.58	5.88	.09	5.03	85.5%
N_2	4.93		4.53		
CO_2	4.21		3.83		
H_2S	trace				

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)

iron sponge to remove CO_2

Actual Total Cost: $363,000 (1964)

Cost Breakdown: (If available)   iron sponge unit $119,000

rest of plant   244,000

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Carson Creek shallow cut high pressure recycling scheme

Gas Type: wet

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	98.7	85.53
Pressure PSI	2000 PSI	34000 to injectors
Temperature °F	70	?
Water Content	saturated	-25°F D.P.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	71.8		82.3		
C_2	11.9		11.9	liquid stream	
C_3	5.8	44.19	2.2	0	0
C_4	3.0	25.32	.2	1.2	5.0%
C_{5+}	4.5	57.28	nil	49.26	86%
N_2	1.8		2.1		
CO_2	1.2		1.3		

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)

molecular sieve unit

Actual Total Cost: $2,011,000 (1963)

Cost Breakdown: (If available) molecular sieve $200,000

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Gilby Texaco
*dry dessicant process

Gas Type: wet

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	35.0	31.9
Pressure PSI	1100	1000
Temperature °F	70	110°F max.

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	81.03		86.65		
C_2	8.00		8.31	liquid stream	
C_3	3.31	23.78	3.23		nil
C_4	1.38	11.58	1.18	1.00	9.0
C_{5+}	1.53	17.09	.18	16.50	96.5
N_2	.42		.45		
CO_2	4.33				
H_2S	trace				

Sulphur Data: n/a

Cost of Plant: (Note any deviations from basic design processes assumed)

dry dessicant plant amine section

Cost Breakdown: (If available)

original plant	$394,000	(1959)
expansion	754,000	(1965)
salt bath heater	18,000	(1966)

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Chevron - Sylvan Lake Gas Type: wet

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	22.123	20.0
Pressure PSI	950	900
Temperature °F	60	?
Water Content	saturated	4 #/MMcf

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	80.25		86.17		
C_2	8.93		9.23	liquid stream	
C_3	3.73	26.5	.59	23.5	88.67
C_4	1.55	12.80	.02	12.0	94
C_{5+}	1.81	20.65	.02	20.46	99
N_2	.77		.84		
CO_2	2.96		3.13		

Sulphur Data: n/a

Actual Total Cost: $1,990,000

Cost Breakdown: (If available)

$ 890,000 (1963)
1,100,000 (1967)

APPENDIX A

GAS PLANT CAPITAL COST ESTIMATE

Plant Name/Area: Lone Pine Gas Type: sour

Gas Inlet & Outlet Data:

	Inlet	Outlet
Volume/Day, MMcf	29.2	23.9
Pressure PSI	1200	900
Temperature ^{o}F	?	?
Water Content	saturated	4 #/MMcf

Gas Analysis:

	MOL%	BBL/MMcf Residue	MOL%	BBL/MMcf Residue	Recovery %
C_1	73.05		86.79		
C_2	3.06		4.31	liquid stream	
C_3	1.42	9.35	1.33	0	0
C_4	1.10	8.35	1.01	1.75	21
C_{5+}	2.21	28.76	.06	27.89	97
N_2	5.59		6.50		
CO_2	3.55		0		
H_2S	10.02		0		

Sulphur Data: Sulphur Production - 110 Recovery Factor - 93.7 (this is high)

Estimated Actual Production - 102 - plant design 112 LT/D.

Actual Total Cost: 1967 - $3,160,000

Cost Breakdown: (If available)

205 foot stack - $130,000 (replacement cost)

APPENDIX B

PRICE INDEX SURVEY

Estimate Number	1963-1964 Estimate %	1964-1965 Estimate %	1965-1966 Estimate %	1966-1967 Estimate %
1	-	-	-	8
2	10	10	10	10
3	10	10	10	10
4	8	8	8	8
5	7	7	7	7
6	9	9	9	9
7	8	8	8	8
8	-	-	10	10
9	8	8	8	8
10	-	10	10	10
11	7	7	7	7
12	10	10	10	10
13	8	8	9	9
14	-	-	-	20
15	8	8	8	8
15	-	12	12	12
17	7	7	7	7
18	-	8	8	8
19	10	10	10	10
20	-	-	10	10
	110	140	161	189
	n = 13	n =16	n = 18	n = 20

Discussion

Each estimate represents the views of a consultant, contractor, or oil company executive who was asked his opinion on the average yearly price increase in gas processing plant capital costs for the four years ending December 31, 1967. Some of those asked would not comment on price

changes for the full four years and there were seven people interviewed who would not venture an opinion on the question asked.

It was decided to calculate each year's average estimate separately instead of attributing a full four years' estimate to those which were given for a lesser period.

Average Yearly Estimated Price Increases

	% Increase	Index No. (1967 Base)
1966-1967 = 189/20 =	9.5%	.913
1965-1966 = 161/18 =	8.9%	.838
1964-1965 = 140/16 =	8.8%	.770
1963-1964 = 110/13 =	8.5%	.709

APPENDIX C

ADDITIONAL ESTIMATES

Plant 1: Sylvan Lake - Hudson's Bay Oil and Gas Ltd.

Operating Parameters:

Q	57 MMcf/D
RC_{5+}	98%
RC_4	nil
RC_3	nil
$H_2S\%$	.71*
$CO_2\%$	2.88

*The amine section was designed to handle up to three per cent hydrogen sulphide at the same inlet flow rate. Therefore, an acid gas content of 5.85% was used for estimation purposes.

Calculation of Estimation Parameters:

$$M_1 = Q\ (1 - a)(1\ RC_{5+} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 57\ (1)\ (98) \times 10^{-2}, \qquad M_1 = 55.9$$

$$M_2 = 57\ (5.88) \qquad M_2 = 335.2$$

$$M_3 = 0$$

Actual Plant Cost:

Date built - 1965, price inflation conversion factor = 1.23

Actual Total Cost	\$2,629,000
Less: Stack Cost Adjustment	230,000*
Cost applicable to study	\$2,399,000

*The stack cost adjustment was based on a verbal estimate of \$380,000 for a three hundred foot brick stack, less the \$150,000 required for the necessary stack today.

APPENDIX C

ADDITIONAL ESTIMATES

Plant 2: Clive-Alix - Hydrogas Ltd.

Operating Parameters:

Q	4.0 MMcf/D
RC_{5+}	92.6%
RC_4	67.1%
RC_3	31.2%
$H_2S\%$	18.17%
$CO_2\%$	2.37%

Calculation of Estimation Parameters:

$$M_1 = Q(1 - a)(1\ RC_{5+} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 4\ (.7946)(9.26 + 1.2 \times 67.1 + 1.6 \times 31.2) \times 10^{-2}$$

$$M_1 = 7.1$$

$$M_2 = 4 \times 20.54 \qquad M_2 = 82.2$$

$$M_3 \text{ (Design)} = 25.0$$

Actual Plant Costs:

This plant is not constructed yet and the estimate is compared with a detailed cost estimate provided by the consulting firm of Barry & Richardson.

Year of construction, 1968 (for price index purposes use a conversion factor of .90)

Costs as estimated (B & R)	\$1,856,000
Less: Compressor & related equipment costs	392,000
Cost applicable to study	\$1,464,000

APPENDIX C

ADDITIONAL ESTIMATES

Plant 3: Kaybob Deep Cut Facilities - Pacific Petroleums Ltd.

Operating Parameters:

Q	12 MMcf/D
RC_{5+}	100%
RC_4	95%
RC_3	70%
$H_2S\%$	nil
$CO_2\%$	nil

Calculation of Estimation Parameters:

$$M_1 = Q\,(1 - a)(1\ RC_{5+} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 12^{(1)}(100 + 1.2 \times 95 + 1.6 \times 70) \times 10^{-2}$$
$$M_1 = 39.1$$

M_2 = nil

M_3 = nil

Actual Plant Cost:

This plant is not constructed yet. The figures below represent an estimate provided by the consulting firm of K.C.M. Engineering Ltd.

Estimate date 1967 (conversion factor = 1.0)

Total plant estimate	\$1,450,000
Less: pipeline costs	160,000
Cost estimate applicable to study	\$1,290,000

APPENDIX C

ADDITIONAL ESTIMATES

Plant 4: Carson Creek Shallow Cut - Mobil Oil Canada Ltd.

Operating Parameters:		
	Q	98.7 MMcf/D
	RC_{5+}	86%
	RC_4	5%
	RC_3	nil
	$H_2S\%$	nil
	$CO_2\%$	1.2%

Calculation of Estimation Parameters:

$$M_1 = Q(1 - a)(1\ RC_{5+} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 98.7^{(1.0)}(86 + 1.2 \times 5) \times 10^{-2}$$

$$M_1 = 90.8$$

M_2 = nil

M_3 = nil

Actual Plant Costs:

Date constructed, 1963 (conversion factor = 1.49)

Total cost of plant	\$2,011,000
Less: molecular sieve cost*	200,000
	\$1,811,000

*The estimated cost of the molecular sieve was deducted as these units are outside the scope of the study.

APPENDIX C

ADDITIONAL ESTIMATES

Plant 5: Wimborne - Mobil Oil Canada Ltd.

Operating Parameters:		
	Q	58 MMcf/D
	RC_{5+}	100%
	RC_4	nil
	RC_3	nil
	$H_2S\%$	19% (calculated)
	$CO_2\%$	nil

Calculation of Estimation Parameters:

$$M_1 = Q(1 - a)(1\, RC_{5+} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 58\ (.81)(100) \times 10^{-2}$$

$$M_1 = 47.0$$

$$M_2 = Q\ A$$
$$= 58 \times 19 \qquad M_2 = 1100^*$$

$$M_3 = .3395 \times 58 \times 19 \qquad M_3 = 373^{**}$$

* and ** - both these estimates are much too large to fit into the scope of this study.

Actual Plant Cost:

Year of construction, 1965 (conversion factor = 1.23)

Total cost of plant	$5,000,000
*Less: sulphur and amine section costs	3,500,000
Liquids section costs, applicable to study	$1,500,000

*Estimated verbally by Mobil personnel to be approximately 70% of the total plant cost.

APPENDIX C

ADDITIONAL ESTIMATES

Plant 6: Berrymoor - Pan American Petroleums Corporation

Operating Parameters:

Q	11.3 MMcf/D
RC_{5+}	98
RC_4	nil
RC_3	nil
$H_2S\%$	nil
$CO_2\%$	nil

Calculation of Estimation Parameters:

$$M_1 = Q(1 - a)(1\ RC_{5+_2} + 1.2\ RC_4 + 1.6\ RC_3) \times 10^{-2}$$
$$= 11.3 \times 98 \times 10^{-2}$$

$M_1 = 11.0$

M_2 = nil

M_3 = nil

Actual Plant Costs:

Construction date, 1965 (conversion factor = 1.23)

Total cost of plant $ 285,000*

*verbal quote